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PROCESS DESIGN MANUAL FOR STRIPPING OF ORGANICS

by

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Contract No. 68-03-3002

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Procedures and correlations for designing and costing stripping towers for the removal of organics from aqueous streams are presented. The emphasis is on practical methods suitable for engineering estimates. The designs cover steam strippers with and without condensers and reflux, as well as air stripping. Steam stripping is treated as an isothermal process and simplified equations for the determination of tower height may be used. Determination of the height of adiabatic air strippers involves a tedious, iterative solution of heat and material balances. A BASIC computer program for carrying out these calculations is provided.

Capital costs are determined, essentially, by estimating the quantity of materials required in conjunction with material costs. Cost factors for a range of materials and installation factors are suggested. Methods for costing ancillary equipment such as heat exchangers, pumps, compressors, and storage vessels are included. It is recommended that vendor quotes be obtained wherever possible. Typical operating costs for energy and maintenance are also given.

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This document has been reviewed in accordance with U.S. Environmental Protection Agency policy and approved for publication. Mention of trade names or commercial products does not constitute endorsement of recommendation for use.

FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory-Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

Stripping is one of several processes that are used commercially for separating organics from water and wastewater. The selection of the optimum process for any given application is dependent on the characteristics of the organics as well as on other factors including cost and the need to recover the separated organics.

In the case of a mixture of organics having different properties, two or more treatments in series may have to be provided. The use of two different processes, for example one with cost dependent on throughput, followed by a polishing step having cost dependent on feed concentration, can result in a lower overall treatment cost. In a wastewater treatment train, stripping is typically the first process that separates dissolved substances. It follows clarification or filtration steps that are used for removal of suspended solids, and may precede polishing steps such as carbon or resin adsorption.

Requests for further information regarding Stripping of Organics should be directed to the Organic and Inorganic Chemicals and Products Branch of the Industrial Pollution Control Division, IERL, Cincinnati.

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ABSTRACT

Procedures and correlations for designing and costing stripping towers for the removal of organics from aqueous streams are presented. The emphasis is on practical methods suitable for engineering estimates. The designs cover steam strippers with and without condensers and reflux, as well as air stripping. Steam stripping is treated as an isothermal process and simplified equations for the determination of tower height may be used. Determination of the height of adiabatic air strippers invovles a tedious, iterative solution of heat and material balances. A BASIC computer program for carrying out these calculations is provided.

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SECTION I

INTRODUCTION

Stripping is one of several processes that are used commercially for separating organics from water and wastewater. The selection of the optimum process for any given application is dependent on the characteristics of the organics as well as on other factors including cost and the need to recover the separated organics. Excellent descriptions of all the important separation processes can be found in the U.S. Environmental Protection Agency's (EPA's) review "Control of Organic Substances in Water and Wastewater".[1] Commonly used separation processes are listed together with some selection criteria in Table 1-1.

In the case of a mixture of organics having different properties, two or more treatments in series may have to be provided. The use of two different processes, for example one with cost dependent on throughput, followed by a polishing step having cost dependent on feed concentration, can result in a lower overall treatment cost.[2] In a wastewater treatment train, stripping is typically the first process that separates dissolved substances. It follows clarification or filtration steps that are used for removal of suspended solids, and may preceed polishing steps such as carbon or resin adsorption.

1.1 STRIPPING APPLICATIONS

The essential characteristic that determines the effectiveness of stripping in separating dissolved organics is the relative volatility or vapor pressure of the organic above

TABLE 1-1

COMMON PROCESSES FOR THE SEPARATION OF ORGANICS FROM WATER

Process	Required Characteristic of Organic	Recovery of Organics	Primary Cost Dependence*
Biochemical Aerobic, Anaerobic	Biodegradable	No, organics are destroyed	Concentration of organics
Solvent extraction	More soluble in solvent than in water	Yes, by fractionating the solvent	Water throughput
Membrane Ultra-filtration, Reverse osmosis	High molecular weight, (ionizable)	Concentrated aqueous stream recovered	Water throughput
Adsorption Carbon, Resin	Adsorbs on selected adsorbent	Not with carbon, possible with resins	Concentration of organics
Stripping	Volatile	Not usually with air; possible with steam stripping	Water throughput

---

*In addition to removal efficiency.

the aqueous phase. It has been shown that at least half of the 186 organics on the EPA's toxic pollutant list are sufficiently volatile to be effectively removed from aqueous waste streams by stripping.[3] Sixty eight of these can be very easily stripped by air at ambient temperatures. Others can be stripped at elevated temperatures with steam. The remaining substances on the list have relatively low vapor pressures and are not easily stripped. The classification of the compounds on the toxic organics list by ease of stripping that is given in Reference 3 is reproduced in the Appendix.

Stripping is emerging as a cost effective alternative for treating a wide range of aqueous streams containing organics. It may be used both as an "in-plant" process for the recovery of organics from relatively concentrated aqueous streams, and as an "end-of-pipe" treatment for removal of dilute and even trace quantities of organics from wastewaters prior to discharge or recycle. Steam stripping is typically used for in-plant separation, whereas air or steam may be used for end-of-pipe treatment, depending on the volatility of the organics and post-treatments provided. In addition, air stripping is being increasingly used for the removal of trihalomethanes (THM's) and trichloroethylene (TCE) from drinking water supplies. References 4-9 contain useful data on pilot and commercial scale strippers.

1.2 PURPOSE

The purpose of this manual is to provide, within a single document, both data and procedures for designing and costing stripping systems for organics separation. A major objective was to develop and summarize simplified and practical engineering procedures of study grade accuracy. The designs and costs obtained are suitable for evaluating the feasibility and viability of stripping relative to other control technologies, and for checking commercial designs. They are not intended for detailed or definitive designs.

1.3 SCOPE

The design procedures cover:

1. Tray and packed towers
2. Air and steam stripping
3. Live and reboil steam
4. Refluxed and non-refluxed steam stripping
5. Isothermal and adiabatic operation
6. Continuous operation
7. Ancillary equipment including heat exchangers

The design and cost procedures are summarized in a stepwise fashion to facilitate their routine use. Procedures for using the simplified analytical equations appropriate to most stripping applications are demonstrated by means of a worked example. A BASIC program suitable for desk-top computers is provided for the case of adiabatic air stripping where the usual simplified equations are less reliable. A comprehensive review of the theory of stripping and the development of the design equations is also included. Although this manual is not intended as a text, the reader may find the background material useful as a refresher course in stripping. We recommend that all users scan the background sections, particularly with reference to the limitations on the procedures and data.

The process design is oriented towards single component relatively dilute systems. Other systems can nevertheless be handled as well. For example, multicomponent systems can be sized by designing for the least volatile organic, and then determining the distribution of the other components separately. Use of the simplified design correlations for concentrated streams may result in errors due to thermal effects and deviations from vapor-liquid equilibrium correlations. The more rigorous design equations may, however, be used without difficulty provided the necessary enthalpy and equilibrium constant data are available.

The design of ancillary systems such as decanters and other vapor handling equipment is specifically not handled. Multiple towers and batch stripping are also not treated. Design procedures for these applications as well as more detailed methods for handling concentrated, multicomponent systems may be found in References 10-12. Maintenance and operational problems are not discussed, nor are controls, instrumentation and civil design covered. The cost of these items is nevertheless included in the budget estimates.

SECTION 2

BACKGROUND TO DESIGN

Stripping operations are characterized by the transfer of mass from one phase to another. In our case, we are specifically interested in the transfer of an organic solute from the water in which it is dissolved to the gas phase.

Stripping occurs in nature. Many pollutants have a half life in a river or lake which is controlled by the rate at which they vaporize into the atmosphere. Bubbling air through the water will greatly enhance the rate of pollutant vaporization and can be used to clean the water. The air that bubbles through the water is called the **stripping medium**. Steam is used as the stripping medium in industrial applications in cases where improved removal is obtained at elevated temperatures.

In industry, stripping is carried out in a **stripping tower** similar to those shown in Figure 2-1. The stripping medium is generally introduced at the base of the tower while the liquid stream, called the **feed**, is introduced at or near the top. When steam is used it may be introduced directly into the base of the tower as **live steam** or may be introduced indirectly as **reboil steam** through a heat exchanger or reboiler. The steam leaving the top of the tower may be condensed and some of the condensate returned to the tower as **reflux**. The tower may be fitted with **trays** or filled with **packing** for enhancing contact between the two phases.

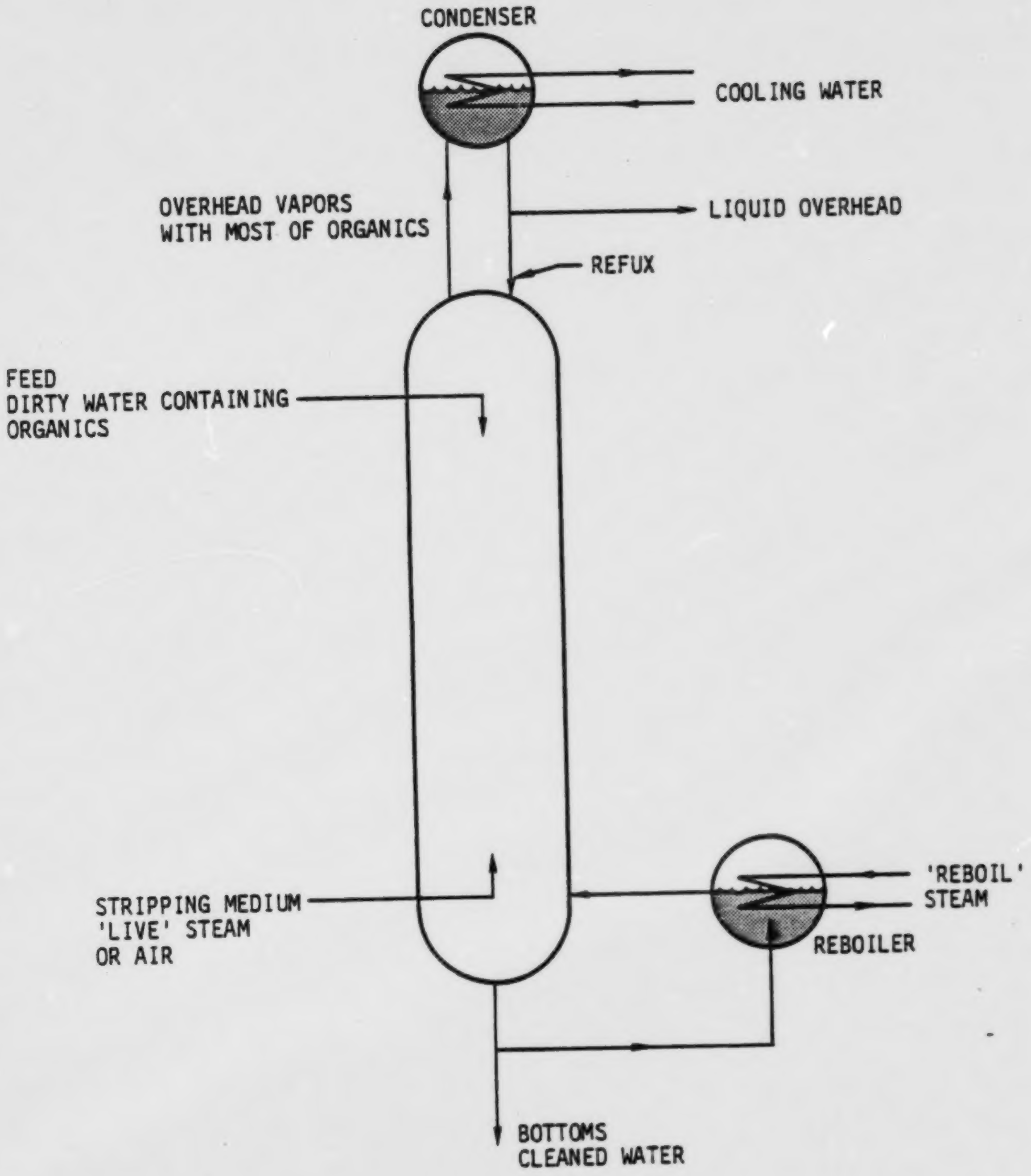

CONDENSER
COOLING WATER
OVERHEAD VAPORS WITH MOST OF ORGANICS
LIQUID OVERHEAD
REFUX
FEED DIRTY WATER CONTAINING ORGANICS
STRIPPING MEDIUM 'LIVE' STEAM OR AIR
'REBOIL' STEAM
REBOILER
BOTTOMS CLEANED WATER

Figure 2-1. Simplified Diagram of a Stripping Tower

The basic design of a stripping tower involves specifying the size of the tower (height and diameter) and the stripping medium flow rate required to strip a given quantity of feed down to a desired purity. More completely, a design requires selecting the type of stripping medium (steam or air), the operating temperature, the type of contacting device (trays or packing), and the combination of tower height and stripping medium flow rate that will achieve the desired separation at minimum overall cost.

The applicable design procedure depends on the selections made amongst the following five types of operation:

1. Batch and continuous. Continuous operation is more effective in separating components of comparable volatility. It provides higher purity of separated products and uses less stripping medium for the same separation. Batch stripping is of less commercial interest for the types of separation under consideration here. We will restrict our design to continuous systems.

2. Mode of flow. Three modes, concurrent, countercurrent and crossflow, as shown in Figure 2-2, are possible. Cocurrent flow is not generally used and is not addressed here. Crossflow operation is often preferred to counterflow because it provides greater transfer efficiency over a wider operating range. Both are discussed.

3. Isothermal or adiabatic. In isothermal operation, the temperature is constant along the length of the tower. If the stripper operates adiabatically, the temperature may change significantly along the tower length. Steam stripping is treated as being isothermal, and the feed is assumed to enter the tower pre-heated to the boiling point. Heat requirements are satisfied by steam condensing at the saturation temperature. Air strippers are treated as being

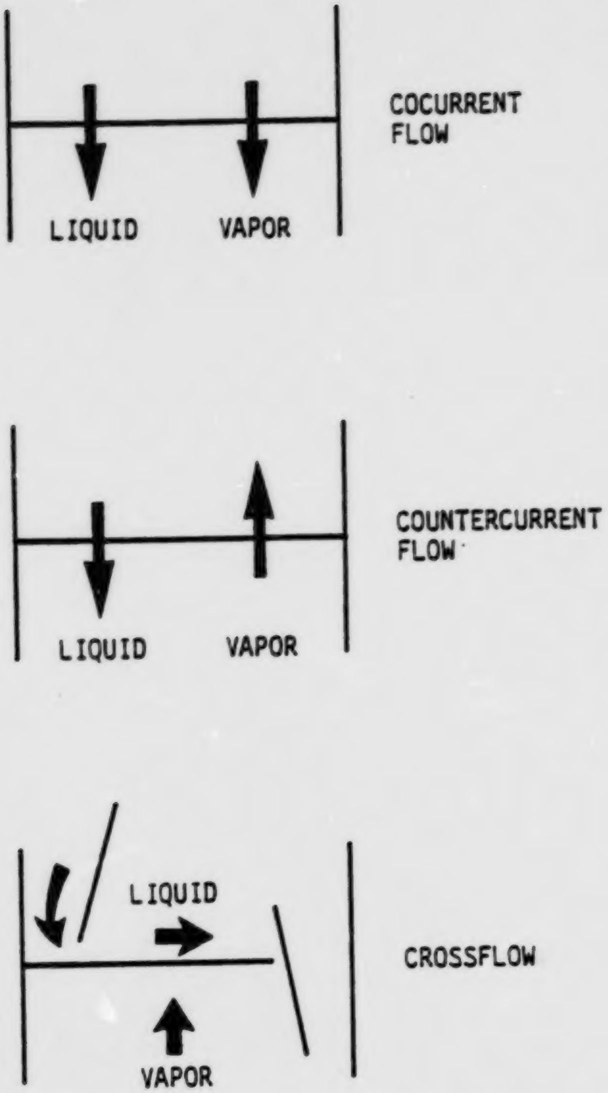

LIQUID
VAPOR
COCURRENT FLOW
LIQUID
VAPOR
COUNTERCURRENT FLOW
LIQUID
VAPOR
CROSSFLOW

Figure 2-2. Modes of Flow in Stripping Equipment

adiabatic. Water usually evaporates into the air, and the water stream is cooled. Saturating the air with water vapor before it enters the tower avoids this problem and can improve separation as the stripping operation then occurs at a higher temperature. This possibility is discussed in Reference 3. The selection between air and steam stripping is essentially dependent on process economics.

4. Reflux. Reflux involves condensing some or all of the vapor leaving the top of the stripping tower, and returning some or all of the condensate back to the tower. Reflux enhances the separation, increases the concentration of the stripped organics in the vapor stream, and is useful if the organics are to be recovered. It is used with steam rather than air stripping. Condensate may be taken to a decanter where an organic phase separates and can be incinerated or recovered. The organic-saturated aqueous phase is returned to the tower as reflux. The design procedures given here for steam strippers are for both refluxed and non-refluxed towers.

5. Mechanism of transfer. This includes differential (staged) contact and integral (continuous) contact. Both are widely used and are discussed in detail under tray and packed towers below.

2.1 TRAY AND PACKED TOWERS

In general, stripping is carried out either in tray towers which provides for staged contact between the liquid and vapor streams, or in packed towers which provides continuous contact between the two phases. The two types of towers are illustrated in Figure 2-3.

In **tray towers** the tower is fitted with regularly spaced trays or plates of the type shown in Figure 2-4. The vapor passes through openings in each tray and contacts the liquid flowing

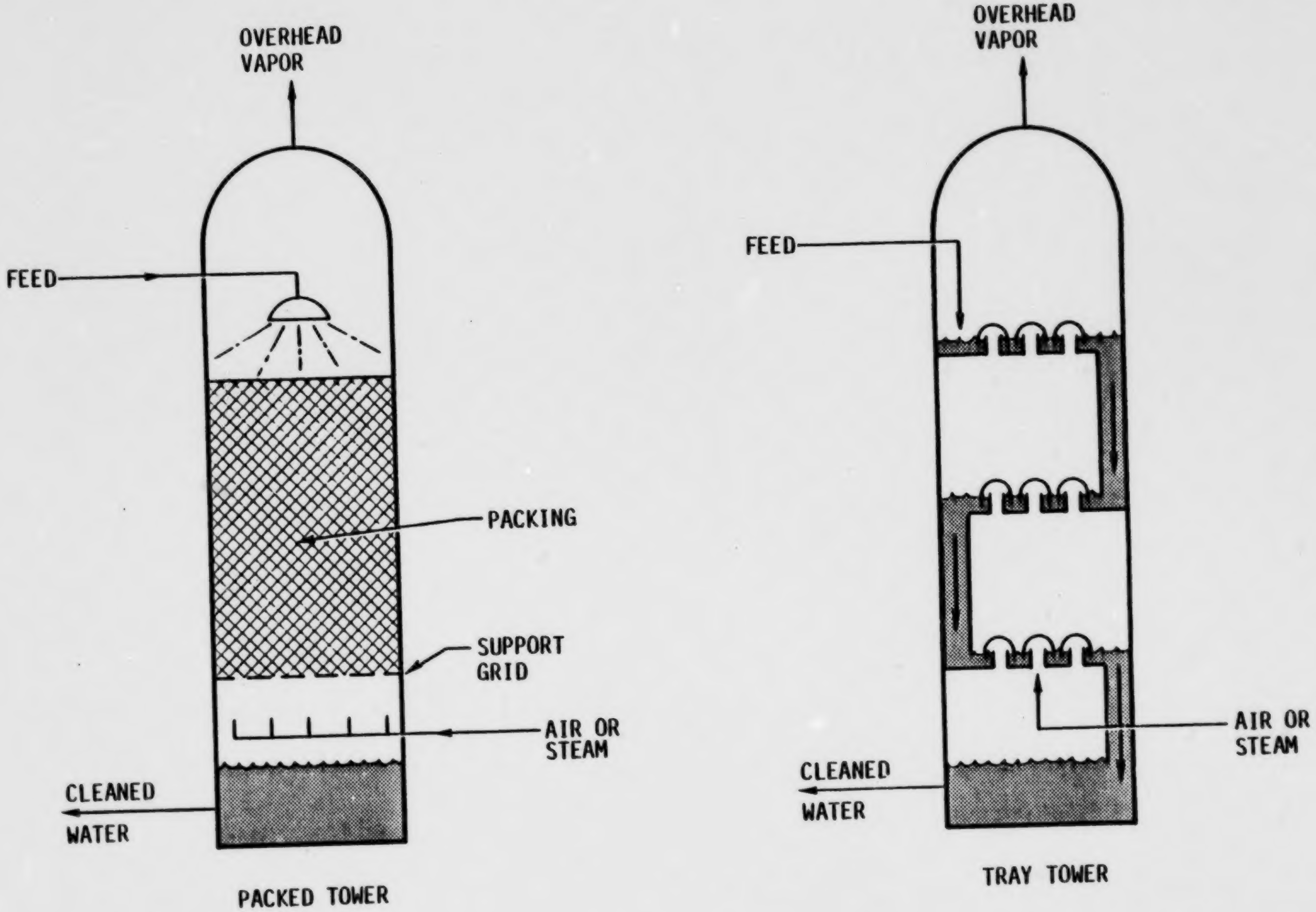

OVERHEAD VAPOR
FEED
PACKING
SUPPORT GRID
AIR OR STEAM
CLEANED WATER
PACKED TOWER
OVERHEAD VAPOR
FEED
AIR OR STEAM
CLEANED WATER
TRAY TOWER

Figure 2-3. Pictorial of Packed and Plate Towers[3]

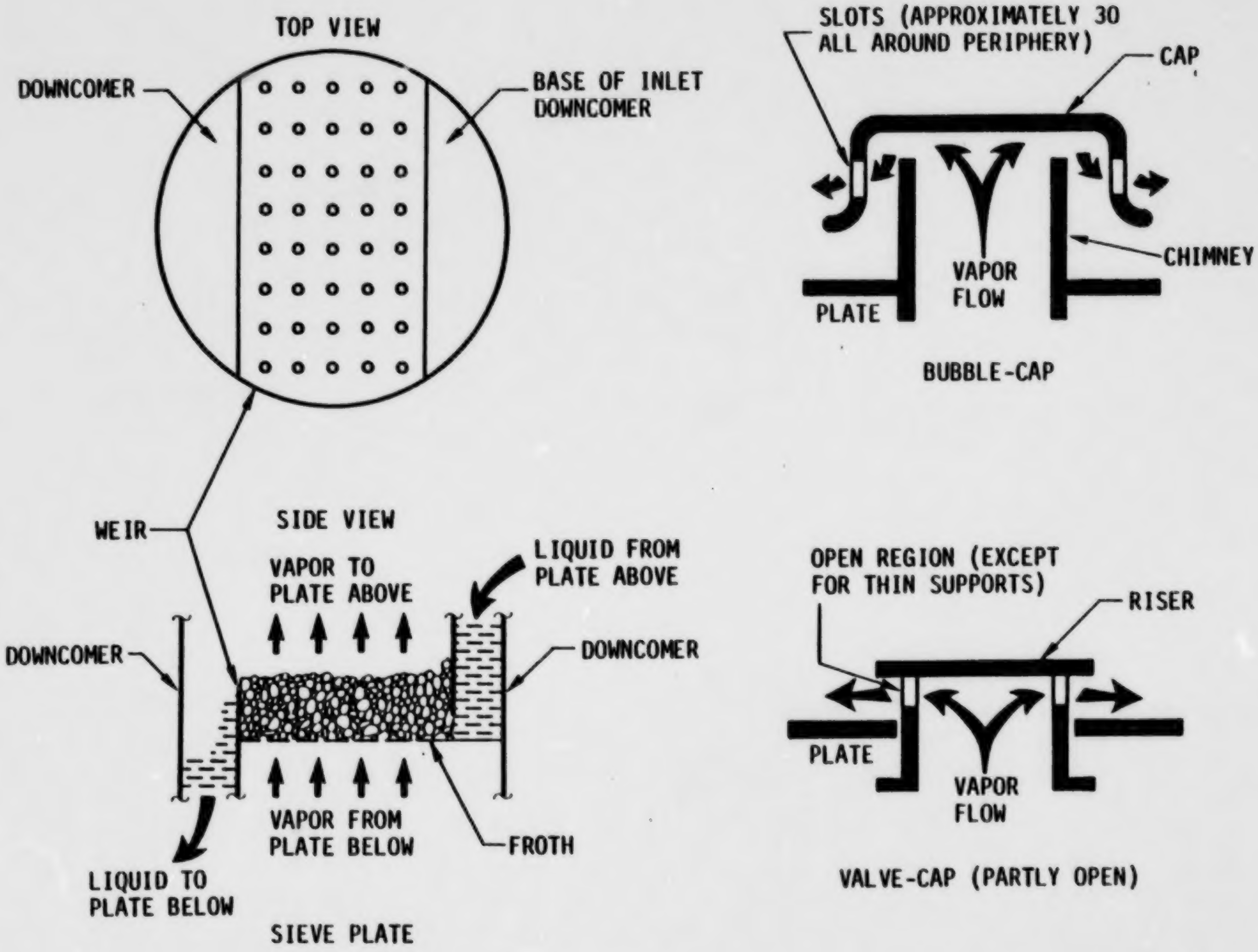

TOP VIEW
DOWNCOMER
BASE OF INLET DOWNCOMER
WEIR
SIDE VIEW
VAPOR TO PLATE ABOVE
LIQUID FROM PLATE ABOVE
DOWNCOMER
DOWNCOMER
VAPOR FROM PLATE BELOW
FROTH
LIQUID TO PLATE BELOW
SIEVE PLATE
SLOTS (APPROXIMATELY 30 ALL AROUND PERIPHERY)
CAP
CHIMNEY
VAPOR FLOW
PLATE
BUBBLE-CAP
OPEN REGION (EXCEPT FOR THIN SUPPORTS)
RISER
PLATE
VAPOR FLOW
VALVE-CAP (PARTLY OPEN)

Figure 2-4. Methods for Dispersing Vapor into Liquid on a Tray[10]

across the tray. A quantity of liquid is retained on each tray by a weir. To reach the next stage, the liquid flows over the weir and through a "downcomer" which provides sufficient volume and a long enough residence time for the liquid to be freed of entrained vapor before entering the next tray. The liquid flows down the tower under the force of gravity, while the vapor flows upward under the force of a pressure drop from tray to tray.

The three common types of trays, namely bubble-cap, sieve and valve trays, are shown in Figure 2-4. Bubble-caps were used widely in the past, but have recently been replaced by sieve and valve types which are less expensive and just as efficient.

Bubble-cap assemblies are in the form of round bell caps and commonly have diameters ranging from 4 to 7 inches. The caps have slots or notches around the lower periphery which aids vapor flow. The slots can be of a saw-tooth type or in the form of punched holes, usually rectangular or triangular. The number of bubble-caps to be used per tray is set by the allowable gas velocity through the slots. Davies[13] has recommended two empirical equations for use in preliminary estimates of minimum and maximum linear slot velocities in distillation towers.

The sieve tray consists of a flat plate perforated with many small holes of 1/8- to 1/2-inch in diameter. The plates are connected with one or more downcomers for handling liquid overflow and may contain weirs and baffles for directing vapor and liquid flows.

Valve trays are sieve trays with lift valves fitted over the holes. Whereas with sieve trays, only the vapor flow prevents liquid from "short-circuiting" and flowing down through the holes, in valve trays, the valve serves as a liquid seal while allowing the passage of vapor. The major advantage of valve trays is that high efficiencies can be maintained over a wider range of operating conditions than with sieve trays. However, valve units are more complex mechanically and are more expensive to fabricate than sieve trays. Design information on valve trays is not readily available.

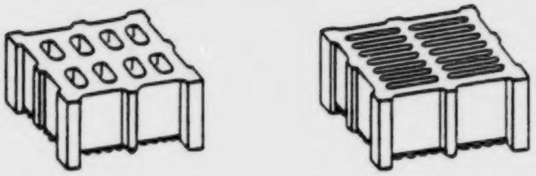
Further information on tray configurations, design and hydraulics is presented by Smith[11].

A **packed tower** is filled with divided solids of the type shown in Figure 2-5. The packing materials may be made of ceramics, metal or plastic, and are shaped to provide a large surface area. The liquid and gas compositions change continuously with height of packing, as opposed to stepwise changes in tray towers.

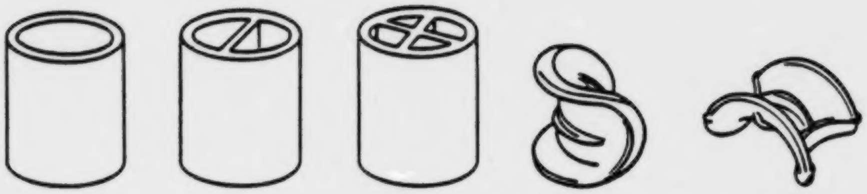
The packed tower is a simple device compared with tray towers. A typical tower consists of a cylindrical shell containing a support tray for the packing material and a liquid-distributor designed to provide effective irrigation of the packing. Devices may be added to the packed bed to provide redistribution of liquid that might channel down the wall. Several beds in series may be used in the same tower shell.

Because of its simplicity, the cost for a packed tower is often considerably less than that for an equivalent tray tower. Packed towers are particularly useful in cases where the pressure drop must be low and where liquid hold-up must be small.

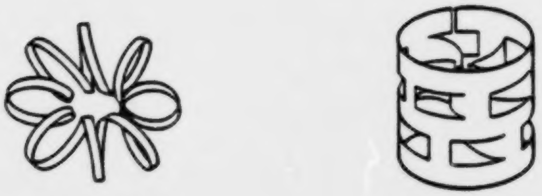
The packings may be either random or stacked. Random packings are simply dumped into the tower. Pall rings, Intalox saddles, Raschig rings and Berl saddles are most often used for random packings in industrial operations. Stacked packings are those in which the individual pieces are arranged in a particular pattern. The larger sizes of Raschig rings (3 inches or larger) may be stacked. Rings are also available with internal spirals which may be stacked one upon the other to provide continuous passages for the gas. Wood grids and drip-point grids are other



DRIP-POINT GRIDS



RASCHIG RING LESSING RING PARTITION RING BERL SADDLE INTALOX SADDLE



TELLERETTE PALL RING

Figure 2-5. Tower Packings[12]

examples of stacked packings. Stacked packings give lower pressure drops for equivalent throughputs than random packings. However, this advantage is gained at the expense of higher initial costs due to the extra installation labor.

2.1.1 Selection Between Tray and Packed Towers

The choice between a tray and a packed tower is often made on the basis of costs, but there are distinct advantages and disadvantages associated with each type of equipment that must also be considered. The following summary of the relative merits of tray and packed towers is based on that given by Peters and Timmerhaus[14]:

1. If the operation involves liquids that contain dispersed solids, use of a tray tower is preferred because the trays are more accessible for cleaning.

2. Tray towers are preferred if interstage cooling is required to remove heats of reaction or solution. Cooling coils can be installed on the trays or the liquid-delivery line from tray to tray can be passed through an external cooler.

3. When large temperature changes are involved, tray towers are often preferred because thermal expansion or contraction of the equipment may crush the packing.

4. Random packed towers are seldom designed with diameters larger than 4 feet, and diameters of commercial tray towers are seldom less than 2 feet.

5. Packed towers prove to be cheaper and easier to construct than tray towers if highly corrosive fluids must be handled. It is easier and cheaper to replace packings periodically than trays.

6. Packed towers are usually preferred for liquids that have a tendency to foam.

7. The liquid holdup is considerably less in packed towers.

8. The pressure drop through packed towers may be less than the pressure drop through tray towers designed for the same duty. This advantage, plus the fact that the packing serves to lessen the possibility of tower-wall collapse, makes packed towers particularly desirable for vacuum operations.

9. Tray towers can operate efficiently over a wider range of liquid flow rates than can packed towers.

Other considerations, while not directly related to the performance of the two types of tower, may nevertheless have a bearing on the selection made. These include:

10. Design information for tray towers is generally more readily available and more reliable than that for packed towers.

11. Because of liquid dispersion difficulties in packed towers, the design of tray towers requires less safety margin when the ratio of liquid to gas flow is low.

12. Reliable design data for packed towers must often be obtained from experiment.

13. The total weight of a dry tray tower is usually less than that of a packed tower designed for the same duty. However, if liquid hold-up during operation is taken into account, both types of towers have about the same weight.

Tables 2-1 and 2-2 provide useful information to help the designer select amongst the various trays and packings available commercially.

TABLE 2-1

RELATIVE PERFORMANCE RATINGS* OF TRAYS AND PACKINGS[10]

	Trays			Packings	
Parameter	Bubble cap	Sieve	Valve	High-void	Normal
Vapor capacity	3	4	4	5	2
Liquid capacity	4	4	4	5	3
Efficiency	3	4	4	5	2
Flexibility	5	3	5	2	2
Pressure drop	3	4	4	5	2
Cost	3	5	4	1	3

---

* 5= excellent; 4= very good; 3= good; 2= fair; 1= poor

TABLE 2-2

SELECTION GUIDE* FOR TOWER INTERNALS[10]

	Trays		Packings	
	Bubble cap	Sieve or valve	Random	Stacked
Pressure, low	1	2	2	3
moderate	2	3	2	1
high	2	3	2	0
High turndown ratio	3	2	1	2
Low liquid flow rates	3	1	1	2
Foaming systems	1	2	3	0
Internal tower cooling	3	2	1	0
Suspended solids	1	2	1	0
Dirty or polymerized solution	1	2	1	0
Multiple feeds or sidestreams	3	3	1	0
High liquid flow rates	1	2	3	0
Small-diameter columns	1	1	3	2
Column diameter 1 to 3m	2	3	2	2
Larger-diameter comuns	1	3	2	1
Corrosive fluids	1	2	3	1
Viscous fluids	1	2	3	0
Low pressure drop	0	1	2	2
Expanded column capacity	0	2	2	3
Low cost	1	2	2	1
Reliability of design	2	3	2	1

---

* 0= do not use; 1= evaluate carefully;
2= usually applicable; 3= best selection

2.2 DESIGN APPROACH

The tower height, liquid and gas flow rates, and the degree of removal of the organics are related by three important concepts, namely:

1. Vapor-liquid equilibrium,
2. Equilibrium or theoretical stages, and
3. Stage efficiency.

Vapor-liquid equilibrium relates the concentrations in the vapor and liquid when these phases are in equilibrium, and an equilibrium stage can be loosely interpreted as the tower height required for the two phases to reach equilibrium. The stage efficiency relates an equilibrium stage to actual conditions in a stripping tower.

Stripping occurs because the dissolved organic, by virtue of its vapor pressure above the solution, tends to vaporize into the stripping medium until its concentrations in the vapor and liquid phases are in equilibrium. In air stripping, some water will also strip (evaporate) into the gas phase until the air becomes saturated. Evaporation or condensation of the liquid or vapor streams may occur in steam stripping as dictated by the thermal effects of the stripping process.

The flow rate of the liquid and gas streams may consequently change due to both mass transfer and thermal effects. These changes are particularly significant for concentrated feeds and complicate the design procedure. When stripping dilute wastewaters with steam or with nearly saturated air, the amount of mass transferred is small relative to the flow, and the liquid and gas rates may be assumed constant. This assumption considerably simplifies the calculations.

Thermal effects associated with heats of vaporization and solution of the components as they pass from the liquid to the gas phase tend to cool the water. In steam stripping, the tower operates at the boiling point of the liquid at the tower pressure. The boiling point is not necessarily constant. This temperature tends to increase down the tower with increasing pressure, but to decrease with decreasing organic concentration. In air stripping, evaporation of water causes the liquid to cool towards the wet bulb temperature as it passes down the tower. Air strippers are said to operate adiabatically. Changes in temperature and flow rate cannot, in general, be neglected for air stripping.

The rate of mass transfer from the liquid to the vapor phase in a stripping system depends upon the difference between the concentration of the solute in the water and the equilibrium vapor concentration at the system temperature. The ease with which a particular compound is stripped or naturally volatilizes depends on the relative volatility, which in turn depends on two properties of the solute - its vapor pressure and its solubility in water. For aqueous mixtures containing low concentrations of organic contaminants, the distribution of a contaminant between the vapor phase and water under equilibrium conditions can often be expressed by **Henry's Law.** That compounds with high vapor pressures are easily stripped is expected. The effect of solubility can also be explained quite simply. A compound which is sparingly soluble in water is not compatible with water molecules. In dilute solutions, each molecule of organic is surrounded by water molecules which want to push the organic molecule away. The apparent vapor pressure of a poorly soluble organic can be thousands of times higher over an aqueous solution than over the pure organic.[3]

2.2.1 Equilibrium Stage

A stage is a separate physical unit into which process fluids are introduced, mixed, separated, and then removed. An equilibrium stage is a stage in which the fluids leaving the stage are in thermodynamic equilibrium. This condition is achieved if the streams entering the stage are mixed thoroughly and for a sufficient length of time for the completion of the required heat and mass transfer processes. An equilibrium stage is sometimes called an ideal or theoretical stage. In tray stripping towers, an equilibrium stage is called a theoretical plate, while in packed towers it is referred to as a transfer unit.

Process systems like stripping towers typically consist of several stages, interconnected so that the materials being processed pass through each stage in turn. The two principal reasons for staging are to increase product purity and to reduce consumption of the separating agent. Multistage systems are called cascades.

The equilibrium stage concept makes it possible to design separation processes despite our poor knowledge of the complex heat and mass transfer operations that occur on an actual stage. Based on this concept, the design of a staged process can be divided into three steps:

1. Determine the equilibrium phase compositions,

2. Calculate the number of equilibrium stages required to accomplish a specified separation, and

3. Estimate the actual number of trays or the height of packing that is equivalent to an equilibrium stage.

The second step involves the relatively straightforward application of equilibrium, material-balance and enthalpy-balance relationships. These are developed in Section 3. In the remainder of this section, we discuss vapor-liquid equilibrium and the concept of stage efficiency in more detail. This is because some empiricism and simplification is required to obtain practical models suitable for design work.

2.3 VAPOR-LIQUID EQUILIBRIUM

Vapor-liquid equilibrium relationships relate the concentration of a species in one phase to the concentration in the other phase when the two phases are in equilibrium. For a detailed description of the theory and estimation of vapor-liquid equilibria, the reader is referred to one of the many texts, for example References 15-17. Here we will provide a brief description suitable for the design procedures presented in this manual.

If we express concentration in terms of **fugacity** the equilibrium "concentrations" in the two phases are identical:

$$f_V = f_L \quad \text{(Equilibrium)} \qquad (2.1)$$

This is the fundamental vapor-liquid equilibrium relationship. By writing the fugacities in terms of commonly used concentration terms, we can convert Eq. (2.1) to a form suitable for design work.

2.3.1 Gas Phase

The mole fraction of a species i in the gas phase is the ratio of the number of moles of the substance n_i to the total number of moles N present:

$$y_i = (n_i/N)_V \qquad (2.2)$$

According to Dalton's law of partial pressures, one can write:

$$p_i = (n_i/N)_V P \qquad (2.3)$$

where p_i is the partial pressure of species i, and
P is the total pressure of the system.

Combining Equations (2.2) and (2.3) we obtain:

$$p_i = y_i \ P \qquad (2.4)$$

Fugacity is defined as being equal to the partial pressure of the species provided the vapor behaves as an ideal gas. Consequently,

$$f_{V,i} = p_i = y_i \ P \quad \text{(Ideal gas)} \qquad (2.5)$$

A correction factor called the fugacity coefficient should be introduced for real gases, but we will assume that pressures are low enough and temperatures high enough for the gas phase to be considered ideal.

2.3.2 Liquid Phase

The mole fraction x_i of a species i in the liquid phase is defined analogously to that in the vapor phase by:

$$x_i = (n_i/N)_L \qquad (2.6)$$

The fugacity is related to the mole fraction by the expression:

$$f_{L,i} = \gamma_i \ x_i \ f^o_{L,i} \qquad (2.7)$$

Two new concepts have been introduced in Eq. (2.7). The standard state fugacity f^o, which can be thought of as the vapor pressure of the pure species, p^*, at the temperature of the solution, again assuming the gas phase is ideal. That is:

$$f_L^o = p^* \tag{2.8}$$

The activity coefficient compensates for non-idealities in the liquid phase. From Eqs. (2.7) & (2.8) we get, for a system containing an ideal gas and a real liquid:

$$f_{L,i} = \gamma_i \, x_i \, p_i^* \tag{2.9}$$

Substitution of Eq. (2.9) for the liquid and Eq. (2.5) for the vapor phase into the basic vapor-liquid equilibrium relationship - Eq. (2.1) - yields an equilibrium relationship which can be applied to design work:

$$y_i = \gamma_i \, x_i \, p_i^*/P \tag{2.10}$$

All terms in this equation are either known or can be easily calculated. The vapor pressure is a function of temperature, and the activity coefficient is a function of both temperature and concentration, so the problem now becomes one of evaluating these parameters at the conditions prevailing in the stripper.

2.3.3 Henry's Law

Many solubility data for gases and volatile organics published in the literature are in terms of their Henry's Law constant, H. This constant is a product of the activity coefficient and pure component vapor pressure:

$$H_i = \gamma_i \, p_i^* \tag{2.11}$$

Substituting in Eq. (2.10) results in Henry's Law:

$$y_i = (H_i/P) \, x_i \tag{2.12}$$

When applying this equation to stripping and distillation calculations, it is customary to write it in terms of a K-factor,

$$y_i = K_i\ x_i \qquad (2.13)$$

where $K = H/P$

Dimensions. Mole fractions are dimensionless, so H has the units of pressure in Eq. (2.12). Other concentration units may be used, in which case the reported H value will have dimensions of pressure divided by concentration. For dilute solutions these H values may be converted to pressure-mole fraction dimensions suitable for use in Eq. (2.12) by multiplying by the following factor:

Concentration Unit	Factor
mg/L or ppm	10^6 M.Wt/18
moles/L	$10^3/18$
moles/m^3	$10^6/18$
mass percent	10^2 M.Wt/18

Temperature and Concentration Dependence. Like the activity coefficient, Henry's Law "constant" is dependent on both the system temperature and the concentration of the substance in solution. As it is unlikely that tabulated H values will be found for the conditions of interest, it is necessary that we be able to estimate these from the available data.

In cases where data is available in the range of interest, plotting or curve fitting and interpolation should be satisfactory. Linearizing the data, as explained below, may improve accuracy.

Many relationships have been proposed between the activity coefficient and concentration.[17,18] At very low concentrations, H may be assumed independent of concentration. At higher concentrations we may extrapolate using the temperature dependence of the activity coefficient. In Reference 3, the van Laar equations for binary mixtures are used, and it is suggested that the data be linearized by plotting

$(x_1 x_2)/(x_1 \ln \gamma_1 + x_2 \ln \gamma_2)$ against the concentration x_1.

Here, subscript 1 refers to the solute and 2 to the solvent (water). The plot has intercept **1/A** and slope **(1/B - 1/A)**. The values of **A** and **B** so obtained may be substituted in the van Laar equation:

$$\text{Log } \gamma_1 = A/[1 + Ax_1/(Bx_2)]^2 \qquad (2.14)$$

to obtain values for the activity coefficient at other concentrations. It must be remembered that **A** and **B** are functions of temperature, so isothermal plots must be used.

Compensating for the effect of temperature on Henry's Law constant is more involved because both the activity coefficient and the vapor pressure vary significantly with temperature. One method of predicting vapor pressure as a function of temperature uses the Antoine equation:

$$\log p^* = A - B/(C + T) \qquad (2.15)$$

A, **B**, and **C** are the Antoine constants (different from van Laar's coefficients), and **T** is the temperature. Reference 16 lists values for these constants for a number of chemicals. A BASIC computer program to find values for **A**, **B**, and **C** from three data points in any (mixed) units is given in the Appendix.

Reed, Prausnitz and Sherwood[17] suggest using:

$$\log \gamma \text{ (constant composition)} = D + E/T \qquad (2.16)$$

where T is the temperature in Kelvin, and
D is a constant that may be assumed equal to zero if only one data point is available.

The correlations suggested above may be used for a wide variety of substances, but are not universally applicable. Refer to Reference 19 for more details.

Data. Values of Henry's Law constant for the organics on the toxic pollutant list that were estimated in Reference 3 are listed in the Appendix. The equations for H are derived from correlations for the vapor pressure and the activity coefficient. In most cases, the activity coefficient was calculated from solubility data. For the few toxic pollutants which are miscible with water, the activity coefficient was estimated from vapor-liquid equilibrium data or from the azeotrope data using the technique of Van Laar. The reader is referred to the original reference for information on these determinations.[3]

2.4 STAGE EFFICIENCY

Stage efficiencies describe the extent to which the performance of an actual stage in the equipment duplicates the performance of an equilibrium stage used in the design calculations. An equilibrium stage normally requires a liquid depth of at least 60 cm, the vapor must be well distributed and in fine bubbles, and flow rates must be low. Such a stage may be used for measuring the Henry's Law constant experimentally but are not practical for trays installed in commercial towers. The relationship between an actual stage or tray and equilibrium stage is normally determined using empirical and simplified mathematical models.

2.4.1. Types of Stage Efficiencies

The **overall tower efficiency**, E_O, is defined as the number of theoretical stages required to produce a given separation divided by the number of trays actually necessary to produce the same separation. Although the overall tower efficiency has no fundamental mass-transfer basis, it is widely used because of its simplicity. The number of actual stages required for a given separation is equal to the number of theoretical stages divided by the overall tower efficiency.

The **Murphree plate efficiency**, E_M, applies to a single stage. It is defined as the ratio of the actual change in average vapor composition accomplished on a tray to the change in average vapor composition if the vapors leaving the tray were in equilibrium with the liquid leaving the tray. The Murphree plate efficiency has a more fundamental basis than the overall tower efficiency, but is less convenient to use because it must be applied to each individual tray.

The **point efficiency**, E_P, is similar to the Murphree plate efficiency, except that the point efficiency applies to a single location on a given tray. This efficiency is of considerable theoretical interest but is seldom used in design because it requires knowledge of the variations in liquid composition across the tray and integration of the point efficiencies over the entire tray.

2.4.2 Tray Towers

The approach to equilibrium depends largely on the effectiveness of mixing the liquid and gas phases on the tray. The overall efficiency has been found to depend on the following factors.[16,28]

1. Viscosity. Efficiency increases significantly as the liquid viscosity decreases. Different authors report different expressions for this dependence but the variation is approximately:

$$\text{Efficiency} \sim (\text{viscosity})^{-n} \qquad (2.17)$$
$$0.7 < n < 0.9$$

   Since the viscosity of water decreases from 0.89 centipoise at 25°C to 0.28 at 100°C there will be a significant increase in efficiency with temperature.

2. Liquid depth. The efficiency increases as the depth of liquid on the tray is increased. The price is an increased pressure drop.

3. Vapor rate. The efficiency is not much dependent on vapor rate up to the point where frothing and entrainment occur. Entrainment causes liquid to be mixed backwards up the tower and the efficiency to decrease.

4. Liquid rate. Since the liquid on bubble-cap and valve trays flows across the vapor, it is possible for there to exist more than one equilibrium stage on a tray. Insofar as an increased liquid rate causes more back mixing in the liquid, the efficiency falls somewhat as the liquid rate increases. The effect is dependent on tray design.

5. Length of liquid path. The length of the liquid path across a tray is an important factor in determining the degree of liquid concentration gradient across the tray. In general, as the length of the liquid path is increased, the overall tower efficiency increases. The effect is usually negligible if the length is less than 5 ft., but increasing the length to 10 to 15 ft. may increase the overall tower efficiency by 20 to 40 percent.

6. Tray spacing. The effect of tray spacing is related to the superficial vapor velocity. If the vapor velocity is greater than the allowable value, too small a tray spacing can cause entrainment carry-over and a decrease in efficiency.

7. Multicomponent separations. In multicomponent separation, the assumption is usually made that the same efficiency applies to all components being separated. This assumption may not be correct for some mixtures where properties of the components (such as liquid diffusivity and Henry's Law constant- see 8 below) are significantly different.

8. Henry's Law Constant. The efficiency may decrease as the Henry's Law constant increases for certain organic compounds. This is a most important consideration in the design of strippers because we are most often concerned with compounds having a high Henry's Law constant. It has been observed that there is not necessarily a direct correlation between a compound's Henry's Law constant and the removal efficiency. This means that for the same conditions, a compound A with Henry's Law constant much higher than that of compound B, may not be much better stripped than compound B. This effect of Henry's Law constant results when the rate of diffusion of the organic through the liquid (rather than across phases) starts to control the overall rate of mass transfer.[3]

9. Other factors. Design details of the tower, such as vapor opening dimensions, tray layout, or the total number of trays can affect the efficiencies.

In general, the three types of trays discussed in Section 2.1 have stage efficiencies in the range of 80 to 90 percent.[14] Sieve and valve trays achieve higher efficiencies than bubble-caps.

2.4.3 Packed Towers

Packing efficiency, which is defined as the ability of a given packing to achieve effective mass transfer between a gas phase and a liquid phase, is inversely related to the height of packing equivalent to one transfer unit (HTU). A packed tower is more efficient the smaller the value of HTU. Apart from geometrical considerations related to the shape and arrangement of the packing, the packing efficiency is affected by the following factors:

Packing efficiency	Factors
Increases	Viscosity decreases Liquid flow rate decreases
Little change	Vapor flow rate alters below the flooding rate
Decreases	Henry's Law constant or relative volatility increases

The height of a transfer unit determined in Section 3 is the actual height and incorporates the factors related to packing efficiency.

SECTION 3

DEVELOPMENT OF DESIGN PROCEDURES

A complete process design for a steam stripping system to recover organics from an aqueous feed could involve specification of all the units shown in Figure 3-1. The system shown in this figure has two stripping towers operating in series, with a common decanter for separating the condensed overheads into organic-rich and organic-lean streams. In the system shown, the feed is preheated by heat interchange with the bottoms.

The design procedures presented here will be broken down into towers, heat exchangers (heat exchanger, reboilers and condensers) and storage vessels (feed tanks, decanters, accumulators, etc.). From these basic components, the designer may synthetize any desired flow sheet. Note that while the design of pumps, flow controllers and instrumentation is not given, the costs for these components are included in the costing procedures given in Section 4.

3.1 TRAY TOWER DESIGN

The design of a tray tower consists of the following determinations:

1. The number of stages theoretically necessary for the required separation,

2. The stage efficiency of the trays relating the theoretical plate to the actual trays,

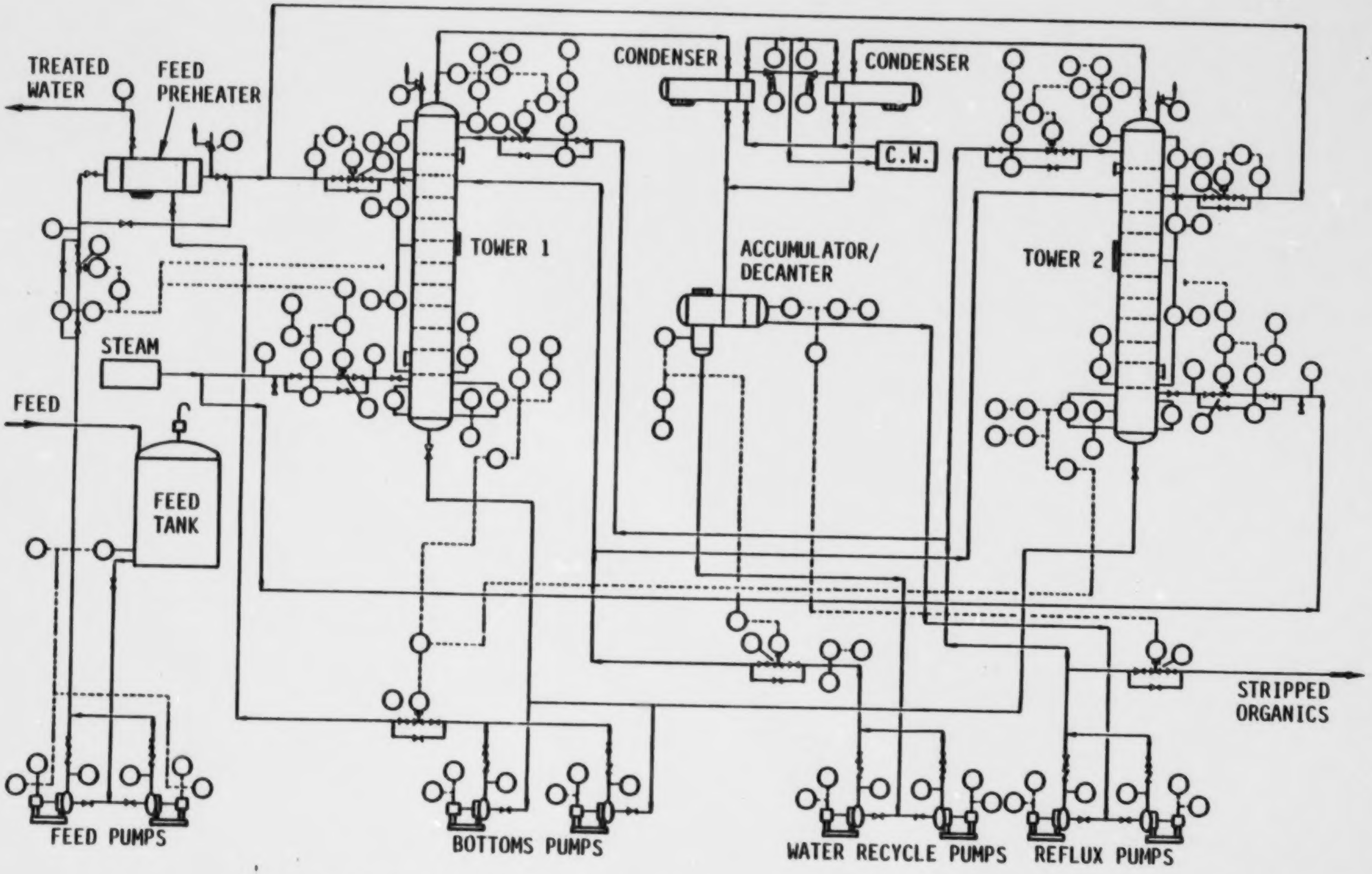

TREATED WATER
FEED PREHEATER
CONDENSER
CONDENSER
C.W.
TOWER 1
ACCUMULATOR/ DECANTER
TOWER 2
STEAM
FEED
FEED TANK
STRIPPED ORGANICS
FEED PUMPS
BOTTOMS PUMPS
WATER RECYCLE PUMPS
REFLUX PUMPS

Figure 3-1. Process Flow Diagram for Steam Stripping

3. The diameter of tower necessary to avoid flooding or excessive entrainment, and

4. The pressure drop across the tower.

3.1.1 Number of Equilibrium Stages

The number of equilibrium stages required to effect a desired separation can be determined from a material balance over the tower. Using the nomenclature in Figure 3-2, an overall material balance for a tower without condenser is:

$$F + A = B + V_1 \tag{3.1}$$

Here, **A** is the total amount of live steam or air introduced at the base of the tower. It is not necessarily equal to V_1, the vapor leaving the top of the tower as transfer of solute and evaporation or condensation takes place along the length of the tower. If reboil steam is used, **A** is zero as no additional material is introduced into the tower. The design calculations are not otherwise different. The effect of using reboil steam is that the bottoms will be more concentrated (for the same steam rate and tower height), but more importantly, clean water (steam) is not introduced into the tower and contaminated. The price for this is the cost of the reboiler.

A material balance for a component with mole fraction x_F, x_B, and y_1 in the feed, bottoms and overhead vapor respectively, becomes:

$$F\ x_F = B\ x_B + V_1\ y_1 \tag{3.2}$$

We can define a fractional removal efficiency, f as the fraction of the solute in the feed that is removed in the overhead:

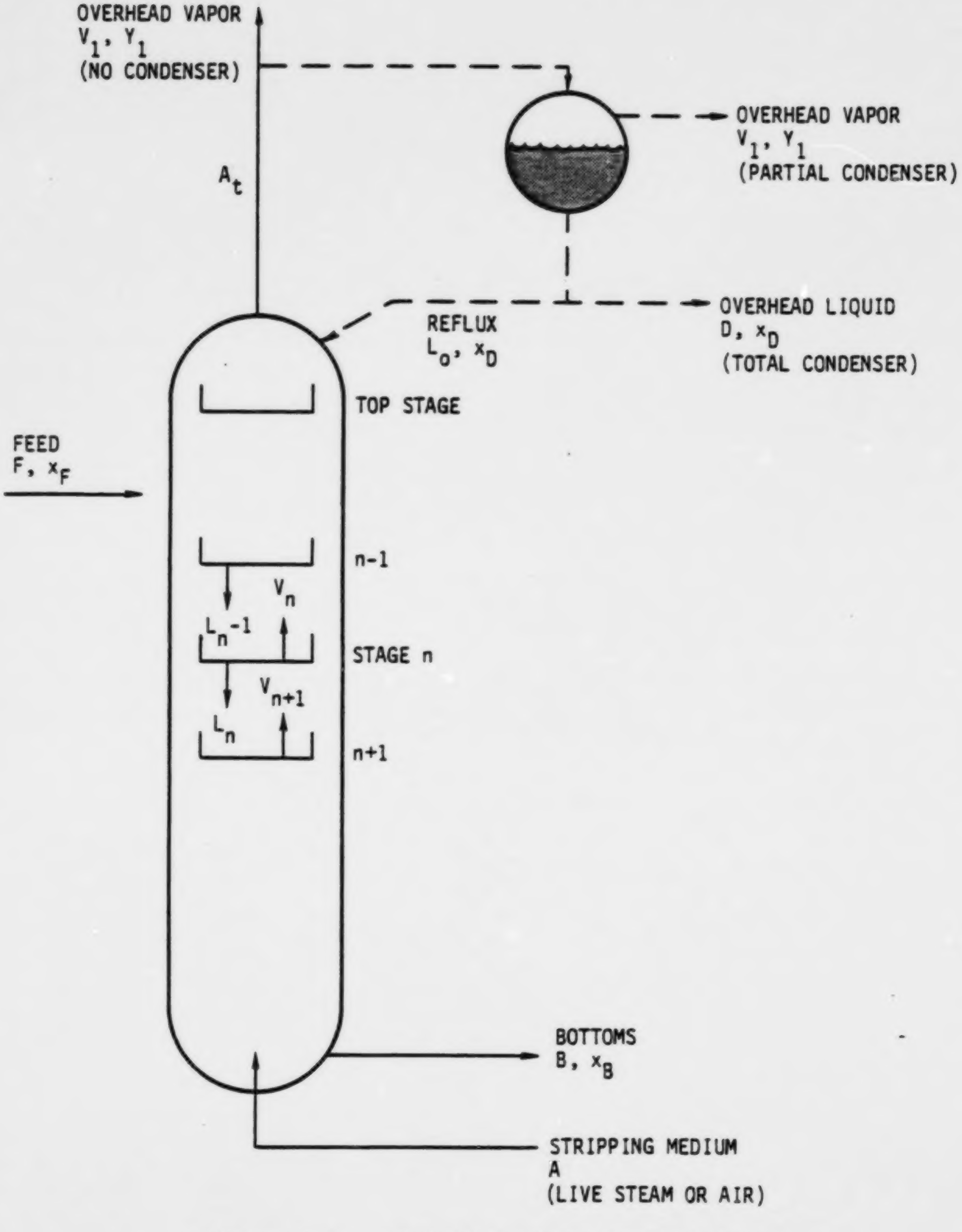

OVERHEAD VAPOR
V_1, Y_1
(NO CONDENSER)
OVERHEAD VAPOR
V_1, Y_1
(PARTIAL CONDENSER)
A_t
REFLUX
L_o, x_D
OVERHEAD LIQUID
D, x_D
(TOTAL CONDENSER)
TOP STAGE
FEED
F, x_F
n-1
V_n
L_n-1
STAGE n
V_{n+1}
L_n
n+1
BOTTOMS
B, x_B
STRIPPING MEDIUM
A
(LIVE STEAM OR AIR)

Figure 3-2. Stripping Tower Nomenclature

$$f = V_1\ y_1/(F\ x_F) \qquad (3.3)$$

In general, we specify the stripping medium rate at the top of the tower A_t (equal to V_1 for the case of no condenser) and the removal efficiency required, and calculate the resulting overhead composition:

$$y_1 = fFx_F/A_t = fFx_F/V_1 \qquad (3.4)$$

The vapor composition on any stage **n** in the tower is obtained by doing material balances from that stage to the top of the tower (See Figure 3-1):

Overall Material Balance

$$F + V_{n+1} = L_n + V_1 \qquad (3.5)$$

Balance for Solute

$$F\ x_F + V_{n+1}\ y_{n+1} = L_n\ x_n + V_1\ y_1 \qquad (3.6)$$

From these two equations, we obtain the vapor composition on stage **n+1** as:

$$y_{n+1} = (V_1\ y_1 + L_n\ x_n - Fx_F)/(V_1 + L_n - F) \quad (n>0) \quad (3.7)$$

The composition $\mathbf{x_n}$ is the liquid composition on the nth tray and is in equilibrium with $\mathbf{y_n}$, the vapor composition leaving that tray. Using Henry's Law we can write:

$$x_n = y_n/K_n \qquad (3.8)$$

Eq.(3.7) now becomes

$$y_{n+1} = (V_1\ y_1 + L_n\ y_n/K_n - Fx_F)/(V_1 + L_n - F) \qquad (3.9)$$

For example, with n=1

$$y_2 = (V_1\ y_1 + L_1\ y_1/K_1 - Fx_F)/(V_1 + L_n - F)$$

Eq. (3.9) can be applied with successive values of n until a vapor composition is obtained that equals or is less than the vapor leaving the bottom tray, y_B. We get the liquid composition leaving the tower, x_B, from Eqs. (3.1) and (3.2) as:

$$x_B = F\ x_F\ (1 - f)/(F + A - V_1)$$

$$= x_F(1 - f) \qquad (3.10)$$

So $$y_B = x_F(1 - f)/K_N \qquad (3.11)$$

If the tower operates isothermally, and K is constant, and the liquid and vapor rates are constant with $L = F = L_n$ and $V = V_n = A_t = V_n$ then Eq. (3.9) simplifies to

$$y_{n+1} = y_1 + (L/VK)y_n - (L/V)x_F \qquad (3.12)$$

The factor (VK/L), called the **stripping factor, S,** is an important parameter in correlating the performance of strippers.

The preceeding equations apply to the case of no condenser. Next we have to consider the use of condensers above the tower. Either a partial condenser or a total condenser may be installed, and some or all of the condensate may be returned to the tower as reflux. The amount of reflux returned to the top of the tower is defined in terms of a reflux ratio **R**. The condenser may be followed by a decanter in which the liquid is separated into organic rich and organic lean fractions. The reflux to the tower may come from the condenser or decanter, or from another source.

The following specific conditions apply to the equations given here for use with condensers:

If a partial condenser is used, it is assumed to be the top equilibrium stage in the tower. All the liquid leaving the partial condenser is assumed to be returned to the tower, so $D = 0$. If a total condenser is used, the reflux composition returned to the tower is the same as the vapor composition leaving the top of the tower, that is $x_D = y_1$. This means that a decanter is not included in the flow sheet. A total condenser is not an equilibrium stage and as the name implies, no vapor leaves the system.

The design equations are summarized in Table 3-1. Note that the generic equations for y_1, y_{n+1} and x_B reduce to those given above for the no condenser case. If reflux is not used, the feed must enter on the top stage. If reflux is used, the feed may enter lower down the tower. Note that F must be zero in Eqs. (23.14) & (3.15) when determining y_{n+1} for stages above the feed tray. The lower the feed tray from the top, the more the overhead vapors are enriched in organics. Two trays above the feed is typical for refluxed strippers. If decanters are included in the flow sheet, do not use the equations in Table 3-1; the simplified equations given later should be used.

Both the stripping medium rate, A_t, and the reflux ratio, R, must be specified when determining the number of stages. If these are not known, a value of A_t equal to 15 percent of the feed for no condenser, and up to 25 percent of the feed for a condenser with reflux may be used initially. If a decanter is used, the entire aqueous part of the overhead is normally returned as reflux. The reader is referred to Reference 12 for a discussion on the minimum stripping medium rates and the effect of reflux on tower performance.

TABLE 3-1

SUMMARY OF STREAM RATES AND COMPOSITIONS*

Variable	No Condenser	Partial Condenser	Total Condenser
Stripping medium	Steam or air	Steam	Steam
Specified parameters			
Stripping medium rate, A_t	V_1	$V_1 + L_o$	$V_1(=L_o+D)$
Reflux ratio, R	0	L_o/V_1	L_o/D
Overhead liquid	L_o, $D = 0$	$D = 0$	L_o, $D>0$
Removal efficiency, f	$V_1y_1/(Fx_F)$	$V_1y_1/(Fx_F)$	$Dx_D/(Fx_F)$
Calculated Parameters			
Composition of overhead vapor	$y_1 = fFx_F(R + 1)/A_t$		(3.13)
Composition of overhead liquid, x_D	-	Eqm. with y_1	$x_D = y_1$
Tower vapor composition n>0	$y_{n+1} = \dfrac{A_ty_1/(R + 1) + L_nx_n - Fx_F}{A_t/(R + 1) + L_n - F}$		(3.14)
or	$y_{n+1} = \dfrac{L_n\ x_n - Fx_F(1 - f)}{A_t/(R + 1) + L_n - F}$		(3.15)
Bottoms composition	$x_B = \dfrac{Fx_F(1 - f)}{F - A - A_t/(R + 1)}$		(3.16)

---

* For design of adiabatic stripping towers without decanters.

Liquid and Vapor Rates. The liquid and vapor rates vary along the tower length due to transfer of the solute and evaporation or concentration of water. To determine these flows, we must combine enthalpy balances with the material balances given previously. Before doing so, it is instructive to first examine under what conditions these rates may be assumed constant.

If the feed is relatively dilute, then both the heat and mass transfer effects of the stripping solute may be assumed negligible. Most organic contaminants are sparingly soluble[3], and we will normally not have to consider concentrated feeds. If more concentrated feeds are encountered, however, the equations can be written on a "solute free" basis. Instead of V_n being the total moles of vapor, it becomes the moles of pure stripping medium, which is constant. Concentrations must then be expressed in terms of mole ratios, (X, Y), that is moles of solute per mole of stripping medium, or per mole of water, instead of mole fractions (x, y). Heat effects must also be accounted for with concentrated feeds. A method for including an enthalpy balance in the solution is discussed under adiabatic air stripping below, and incorporated in the BASIC design program in the Appendix.

In **steam stripping**, the feed can be assumed to be a saturated liquid and each equilibrium stage taken to be at the saturated water vapor temperature. Because the pressure increases from the top to the bottom of the tower, so does the saturation temperature. Consequently, some steam must condense on each stage to supply the heat necessary to raise the liquid temperature. For low pressure drops, this condensation may be neglected and the column assumed to operate isothermally with L_n and V_n constant.

In **air stripping**, water will generally evaporate to saturate the air, and the liquid stream will cool down towards the wet bulb temperature. A constant water rate equal to the feed rate less the water vapor transferred to the air may be assumed for rough estimates. If computers are not available and simplified isothermal calculations must be made, then it is suggested that a temperature which is the arithmetic mean of the feedwater temperature and air wet bulb temperature be used[3].

Because of the great difference in tower performance that can result between summer and winter when stripping with air, it has been suggested that the cooling effect be eliminated by saturating the entering air with steam.[3] If the air is simultaneously heated to near the liquid feed temperature, then the simplified design equations for constant flows and isothermal conditions should satisfactorily approximate the tower performance.

Simplified Equations for Steam Stripping

Making the assumption of constant liquid and vapor rates along with isothermal conditions considerably simplifies the calculational procedure. Because the Henry's Law constant is now also invariant, an analytical solution can be obtained to the design equations. The solution is summarized in Table 3-2. These equations, based on a derivation by Kremser[12], are applicable to most steam stripping operations.

Adiabatic Stripping with Air

In general, the simplified equations will not be suitable for air stripping. The following procedure should be used instead. The equations are derived for strippers without condensers treating dilute wastewaters (the usual application for air stripping). The equations in Table 3-1 for no condenser apply. The vapor rate is constant if we use pure (water- and solute-free) air for **V**. We must express vapor concentrations as mole ratios **Y**, where

$$Y = y/(1 - y) \qquad (3.19)$$

TABLE 3-2

DESIGN EQUATIONS* FOR TRAY TOWERS[20]

$$f = (S^{N+1} - S)/[S^{N+1} - S + k(S - 1)] \qquad (3.17)$$

$$N = \ln[S + f\,k(S - 1)/(1 - f)]/\ln S - 1 \qquad (3.18)$$

where f = fractional removal efficiency of a solute

S = stripping factor $(= KV/L) > 1)$

N = number of theoretical stages

$k = (1 + R'\,\gamma_D/\gamma_S)/(R' + 1)$

$R' = R\,\{1 + [C_p(t_b - t_R)/L]\}$

R = reflux ratio

t_R = reflux temperature

t_b = boiling point of the aqueous reflux

L = latent heat of steam

C_p = specific heat of reflux

γ_D = activity coefficient of pollutant in the organic phase (= 1.0 for pure organic)

γ_S = activity coefficient of pollutant in the aqueous phase[3,20]

---

* For isothermal operation, constant flow rates, with or without reflux and decanter.

The temperature and the amount of water evaporated for each stage can be determined from an enthalpy balance. An enthalpy balance over the top portion of the tower is:

$$F\ h_f + V\ i_{n+1} = V\ i_1 + L_n\ h_n \qquad (3.20)$$

Here h is the water enthalpy in calories/mole,
i is the humid air enthalpy in calories/mole dry air,
V is the pure, dry air rate in moles.

The amount of water transferred on stage n is given by

$$W_n = (H_n - H_{n+1})\ V \qquad \text{moles water} \qquad (3.21)$$

where H is the humidity, moles water vapor per mole dry air.

The above equations can be solved together with the following correlations for enthalpy and humidity.[3]

Water enthalpy:

$$h = 18\ t \qquad \text{cals/mole water} \qquad (3.22)$$

Enthalpy of humid air

$$i = 6.96\ t + H(10{,}750 + 8.1\ t) \quad \text{cals/mole dry air} \qquad (3.23)$$

Mole fraction water vapor in air:

$$y = (1/P)\exp[21.2 - 5921/T - 6.98\times10^{-3}T] \qquad (3.24)$$

Humidity

$$H = y/(1 - y) \quad \text{moles water vapor/mole air} \qquad (3.25)$$

Where t is the temperature in °C, T is in Kelvin, and
P is pressure in psi

We must know the top temperature to initiate the calculations. A temperature which is the mean of the feed temperature and wet bulb temperature of the entering air may be used as a first guess. The calculations then proceed stagewise down the tower. First the temperature and water rate from the top stage ($n = 1$) are calcutated using the above equations. Next the equilibrium concentrations are calculated using the Henry's Law constant at the calculated temperature and the equations in Table 3-1. The procedure is then repeated for the second stage and so on until the bottom stage is reached. If the feed air temperature does not match the calculated temperature for the air entering the bottom stage, a new top temperature must be guessed. A computer program for these calculations is given in the Appendix.

Some results obtained using the computer program are shown in Figure 3-3. These data illustrate the effect of the temperature of the liquid and air feed, as well as the air humidity, on stripping tower design.

3.1.2 Tower Diameter and Height

The tower diameter and consequently its cross section must be sufficiently large to handle the gas and liquid at velocities which will not cause flooding or excessive entrainment. The diameter can be obtained from a determination of the maximum allowable vapor velocity. The following equation applicable for any specific location in the tower may be used[21,22]:

$$v_m = K_v(\sigma/20)^{0.2}[(\rho_L - \rho_G)/\rho_G]^{0.5} \qquad (3.26)$$

where v_m = max. allowable vapor velocity based on cross-sectional area of empty tower, ft/s

K_v = empirical constant, ft/s

ρ_L, ρ_G = density of liquid and gas respectively, and

σ = liquid surface tension, dyne/cm

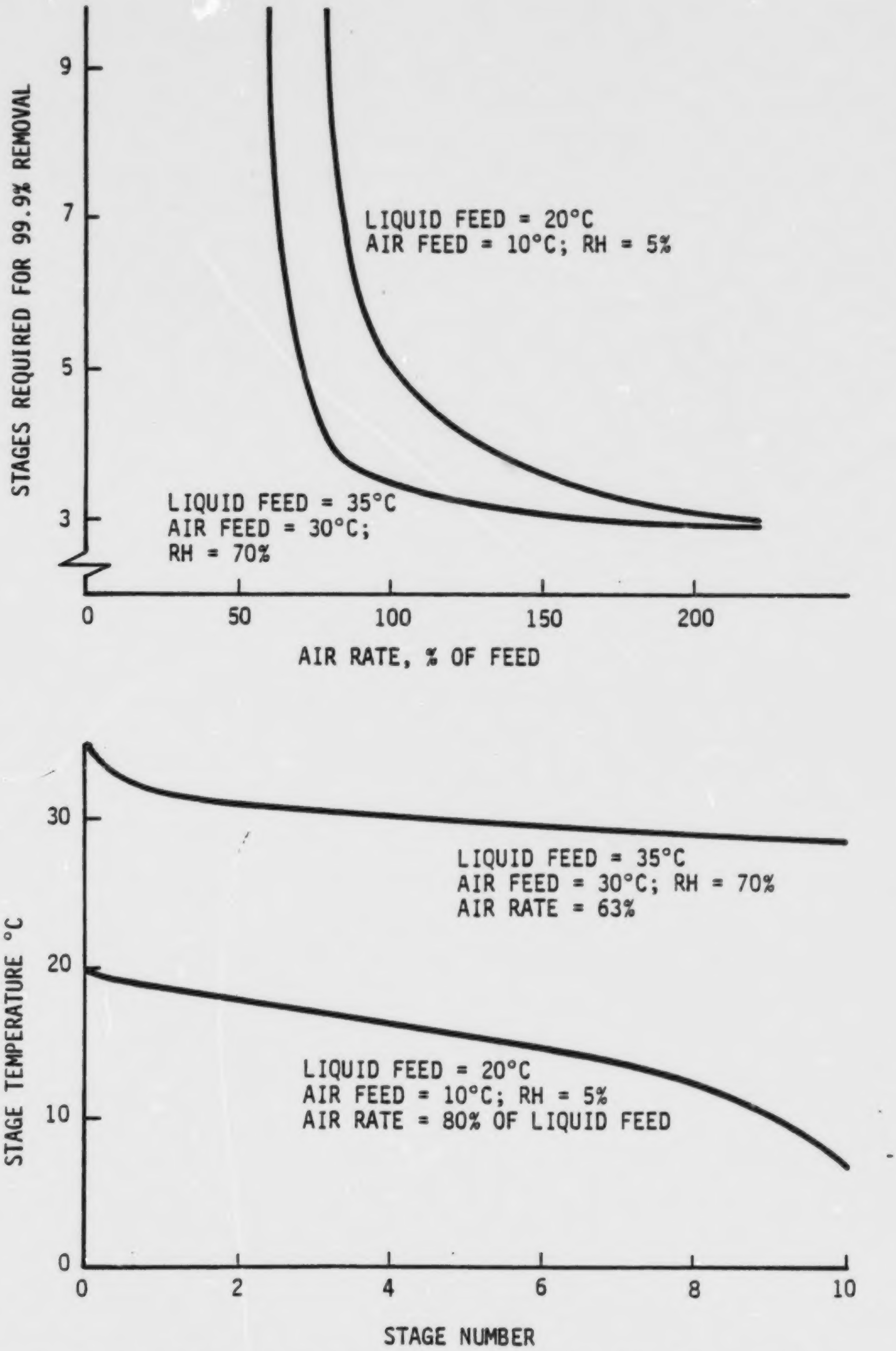


Figure 3-3. Effect of Temperature on Air Stripper Design

The constant K_v is dependent mainly on the tray spacing but also on the depth of liquid on the tray, the ratio of liquid to gas flow rate, the density of the gas and liquid, and the geometry of the tray components. Values for the empirical constant can be obtained using the equations given below.[12]

For bubble-cap trays:

$$K_V = a \log(1/c) + b \tag{3.27}$$

For perforated trays:

$$K_V = [5(A_h/A_a) + 0.5][a \log(1/c) + b] \tag{3.28}$$

where A_a = Active area of the tray which is the total cross-sectional area of tray minus the area of the two downcomers (ft^2).

A_h = Hole (perforation) area per tray.

$c = (L/G)(\rho_G/\rho_L)^{0.5}$

The values of **a** and **b** and the recommended values for tray spacing are given in Table 3-3. A tentative value of 0.13 may be used for A_h/A_a in the absence of other data.

Although the allowable velocity given by these methods is conservative for many types of operations, one can employ velocities in the range of 65 to 85 percent of V_m to ensure the tower will be operable. The volumetric flow rate of the vapor is

$$Q = G/(3600\ \rho_G) \quad ft^3/s$$

and the cross-sectional area required to give a velocity v ft/s is

$$A = 1.1\ Q/v \quad ft^2 \tag{3.29}$$

The factor 1.1 is to allow for area occupied by the downcomers.

TABLE 3-3

RECOMMENDED VALUES FOR TRAY SPACING AND FLOODING CONSTANTS FOR TRAY TOWERS[12]

Tray spacing, t

Tower diameter, ft	Tray spacing, inches
4 or less	18-20
4-10	24
10-12	30
12-24	36

Flooding constants in terms of tray spacing, t

Bubble-cap trays

Value of c	a	b
0.01-0.03	Use values for 0.03	
0.03-0.2	0.0041t + 0.0135	0.0047t + 0.068
0.2-1.0	0.0068t + 0.049	0.0028t + 0.044

Perforated trays

0.01-0.1	Use values for 0.1	
0.1-1.0	0.006t + 0.04	0.0025t+0.05

If the vapor rate changes over the length of the tower, the theoretical diameter based on the available vapor velocity will also change. Vapor rates at different heights of the tower (usually bottom, middle and top) can be used to estimate diameters. Occasionally, two different diameters are used for different sections of one tower. When the variation in flow is considerable, and especially when expensive alloys are used in construction, use of two diameters may be considered. Usually a single diameter based on the tower location where allowable velocity requires the largest diameter, is used.

Once the diameter has been determined, the recommended tray spacing from Table 3-3 together with the number of trays as determined in the next Section may be used to determine the height of the tray section of the tower.

3.1.3 Stage Efficiency

The number of theoretical stages determined must be divided by the overall tower efficiency E_O to obtain the number of actual stages required for the desired removal. There are several ways to estimate the tower efficiency. Most of them are empirical and developed for a particular application. The method presented here resulted from a research program sponsored by the American Institute of Chemical Engineers. Most of the work was done with bubble-cap trays, but the few data together with subsequent information indicate that the empirical expressions represent the performance of sieve trays as well. The method utilizes the concept of transfer units which are discussed more fully in Section 3.2. The procedure is presented without discussion in the Appendix. The equations are given in a simplified form that are valid for typical conditions; the reader is referred to Reference 23 for further details.

3.1.4. Pressure Drop

As the vapor passes through the tray tower assembly, its pressure decreases due to:

1. Pressure drop through the contactor assembly. This includes

   Contraction loss as the vapor enters the tray openings;
   Friction in the tray opening and in the annular space if a bubblecap unit is used;
   Friction due to change in direction of vapor flow for bubblecap and valve units;
   Friction as the vapor through the slots for a bubblecap unit.

2. Pressure drop due to the static liquid head above slots, sieve openings, or valve openings.

A detailed method of calculating pressure drops for individual steps is given by Peters and Timmerhaus[14]. For a bubblecap tower, the following pressure drops per tray are considered typical, and may be used for sieve or valve trays if other data are not available.

Total pressure	Pressure drop per tray
30 mm Hg	3 mm Hg or less
1 atm	0.07-0.12 psi
300 psi	0.15 psi

3.2 PACKED TOWER DESIGN

The design of a packed tower consists of the following steps:

1. Selection of the packing type and size.
2. Calculation of the total height of packing required.
3. Estimation of the tower diameter to avoid flooding.
4. Estimation of the pressure drop.

3.2.1 Selection of Packing

The types of packing available are described in Section 2.1, and a summary of their important characteristics is given in Table 3-4. There are many packing types on the market and it is frequently advantageous to compare the characteristics of more than one packing type and size before settling on a final design. The following properties are representative of a satisfactory packing:

1. Low pressure drop and high capacity. A large free cross-sectional area should be available between the packing particles to give low pressure drops and high fluid rates.

2. Large active surface area per unit volume. To give high efficiencies, the packing must provide a large amount of contact area between the two fluid phases.

3. Low weight and low liquid holdup. The total weight of the tower and the resultant foundation load is low if the weight of the packing and the liquid holdup in the tower are low. The amount of liquid holdup, however, must be sufficiently great to retain an effective driving force for mass transfer.

4. Miscellaneous. High durability, high corrosion resistance, low side thrust on tower walls and low cost must all be considered in the selection of a packing.

3.2.2 Tower Diameter

The flow rates through the packed bed must first of all be high enough to ensure that the packing is "loaded", that is that the packing surface is continuously wetted and good distribution of the liquid and vapor flows occur. However, if the flow rates are too high, the packing will "flood" with rapid increase in pressure drop and loss of performance. The tower diameter is generally selected to be as small as possible for economic reasons, but large enough to ensure that flooding will not occur under anticipated operating conditions.

Flooding and pressure drop data are provided by packing manufacturers. Figure 3-4 along with Table 3-4 may be used for a variety of packings.

The method of calculating the tower diameter using Figure 3-4 can be summarized as follows:

1. Calculate the parameter $(L/G)(\rho_G/\rho_L)^{0.5}$. Note that L and G are in mass units.

2. At flooding, find from Figure 3-4 the corresponding value of the parameter (with μ_L in centipoise):

$$(G')^2 C_f(\mu_L)^{0.2}/(g_c \rho_G \rho_L)$$

3. Calculate G', the superficial mass velocity, (G/A).

4. Reduce this flow by a flood safety factor, called the flood ratio. A flood ratio of 0.7 to 0.8 is typically used but it may be lower if dictated by pressure drop limitations.

5. The tower cross-section is the entering flow rate of the gas, G, in lb/hr divided by G' in lb/(hr.ft^2).

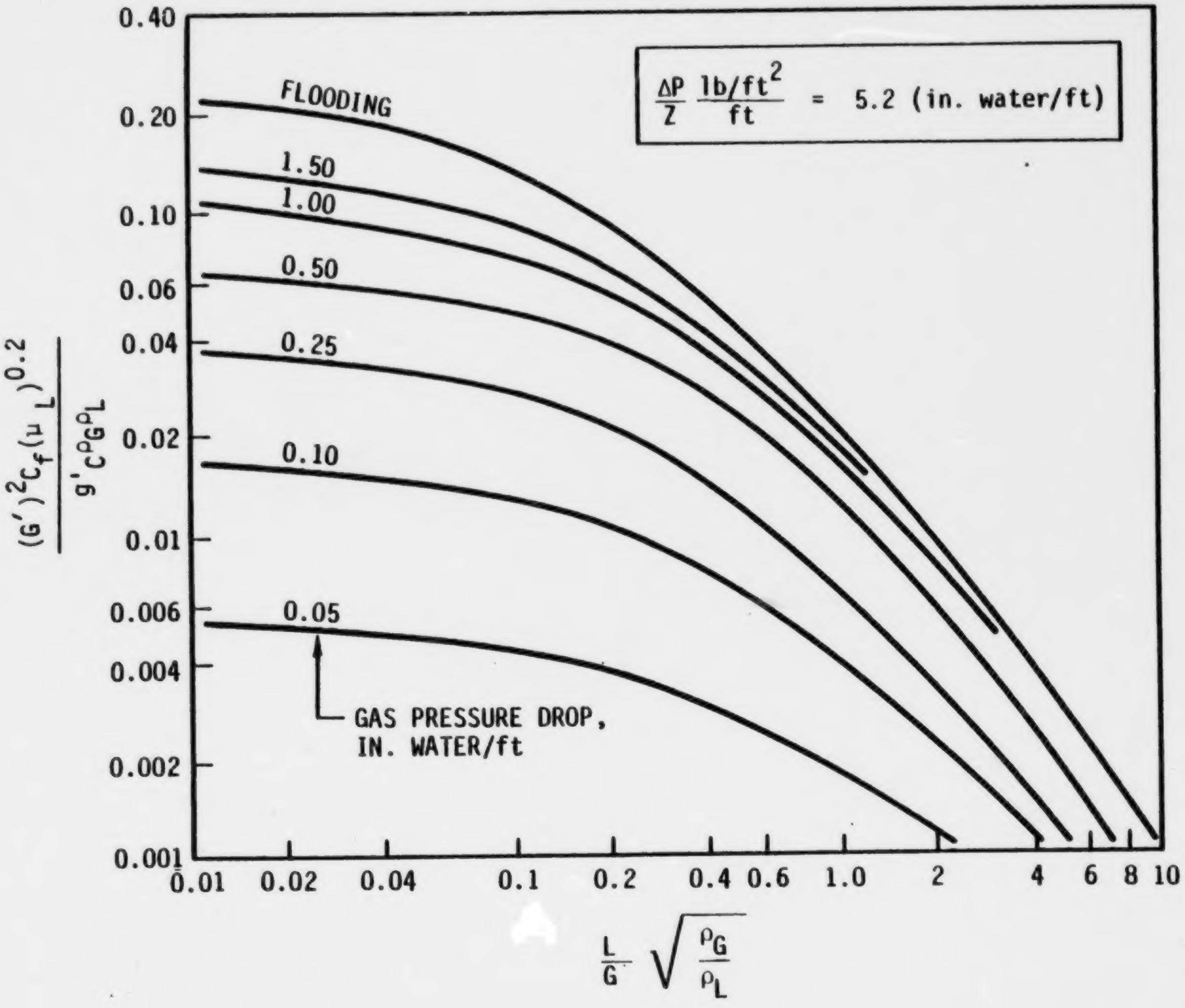


Figure 3-4. Flooding and Pressure Drop in Packed Towers[12]

TABLE 3-4

CHARACTERISTICS OF COMMERCIAL PACKINGS[12]

Data for Wet-Dumped (Random) Packings*

	Nominal size,inches					
Packing	1/2	5/8	3/4	1	1-1/2	2
Raschig rings						
Ceramic						
C_f	640	380	255	160	95	65
ε	0.63	0.68	0.73	0.73	0.71	0.74
a_p	111	100	80	58	38	28
Metal						
C_f	340	290	230	145	82	57
ε	0.73		0.78	0.85	0.90	0.92
a_p	118		71.8	56.7	41.2	31.4
Pall rings						
Plastic						
C_f		97		52	32	25
ε		0.88		0.90	0.905	0.91
a_p		110		63	39	31
Metal						
C_f		71		48	28	20
ε		0.90		0.94	0.95	0.96
a_p		131.2		66.3	48.1	36.6
Intalox saddles						
Ceramic						
C_f	265		130	98	52	40
ε	0.78		0.77	0.78	0.81	0.79
a_p	190		102	78	59.5	36
Berl saddles						
Ceramic						
C_f	380		170	110	65	45
ε	0.63		0.66	0.69	0.75	0.72
a_p	142		82	76	44	32

---

*C_f= packing factor, ft^{-1}; ε = voidage;
a_p= superficial area, ft^2/ft^3

3.2.3 Height of Packing

The concept of a transfer unit is useful for interpreting and correlating mass tansfer data. It is based on the idea of dividing the packed section into a number of contact units called transfer units. Although transfer units are, in fact, equilibrium stages, the derivation of the design equations differs from that given for tray packings. In addition to the texts,$^{10-12}$ the reader is referred to Reference 24 for a good description of the design of packed air strippers.

The depth of packing required by one transfer unit is called the height of a transfer unit (HTU). The total height of the packed section is given by the relation:

$$Z = (HTU)(NTU) \tag{3.30}$$

where Z = packed height
HTU = Height of a transfer unit
NTU = Number of transfer units

The NTU characterizes the difficulty of stripping a compound to a desired level whereas the HTU characterizes the efficiency of mass transfer. The usual order of magnitude of HTU is 0.5 to 5 ft. The actual value should be determined from pilot studies for the system under consideration, but data supplied by packing manufacturers are also useful. In the absence of data, mass transfer correlations or models from the literature can be used for feasibility analyses.

In this section, the equations are presented without derivation. Much of the discussion in Section 3.1 on tray towers is relevant, especially with reference to the constraints related to temperature and flow rates. The design equations given here are for isothermal towers and constant flow rates.

Height of a Transfer Unit (HTU)

HTU can be estimated in several ways.[24-26] Most of the methods use some kind of mass transfer model. Good experimental data are required to relate packing performance to mass transfer models. These models must account for resistance to transfer in both vapor and liquid phases. To accomplish this, the transfer height is taken to be made up of the heights required for transfer through the liquid and gas phases:

$$\mathrm{HTU} = (\mathrm{HTU})_V + S\,(\mathrm{HTU})_L \qquad (3.31)$$

$(\mathrm{HTU})_V$ and **$(\mathrm{HTU})_L$** can be calculated using the equations given in Table 3-5, which were developed by Bolles and Fair.[26] Note that the value for HTU is itself a function of the tower height, so the height determination is an iterative procedure.

Number of Transfer Units (NTU)

The NTU is defined in terms of an integral of the feed **(x_F)** and bottoms **(x_B)** concentrations. This integral equation can be solved analytically for isothermal operation, constant flow rates of dilute solutions, and solutes obeying Henry's Law. The resulting equation to determine NTU for packed towers with and without reflux is:[20]

$$\mathrm{NTU} = S/(S - 1)\ln[(k(S - 1)/\{S(1 - f)\} + 1 + k(1 - S)/S] \qquad (3.34)$$

The nomenclature is defined in Table 3-2.

TABLE 3-5

DETERMINATION OF HTU FOR PACKED TOWERS[26]

$$(HTU)_V = \Psi d^m (Z/10)^{1/3} (Sc_V)^{0.5} / (3600\ G'\ f_\mu\ f_\rho\ f_\sigma)^n \qquad (3.32)$$

where $(HTU)_V$ = Height of a vapor phase transfer unit, ft

Z = Total packing height, ft

Ψ = Vapor phase packing parameter. Determined from Figure 3-5

G' = Vapor flux, lb/(hr.ft^2)

d = The lesser of tower diameter or 2 ft

m = 1.24 (rings), 1.11 (saddles)

n = 0.6 (rings), 0.5 (saddles)

$f_\mu = (\mu_L/\mu_W)^{0.16}$

$f_\rho = (\rho_L/\rho_W)^{-1.25}$

$f_\sigma = (\sigma_L/\sigma_W)^{-0.8}$

$Sc_V = \mu_V/(\rho_G D_V)$

$$(HTU)_L = \phi C_{fL} (Z/10)^{0.15} (Sc_L)^{0.5} \qquad (3.33)$$

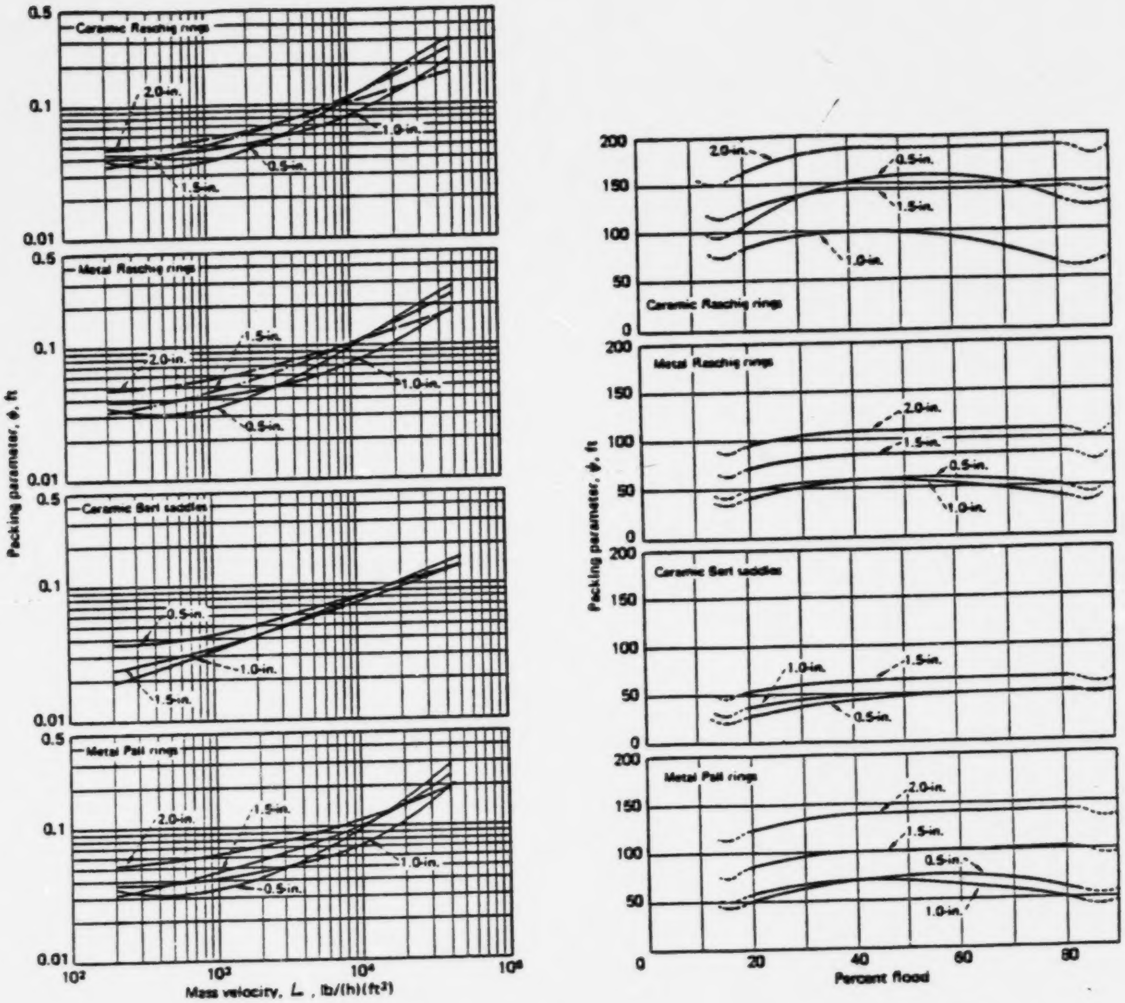
where $(HTU)_L$ = Height of a liquid-phase transfer unit, ft.

ϕ = Liquid phase packing parameter. Determine from Figure 3-5.

$Sc_L = \mu_L/(\rho_L D_L)$

C_{fL} = Vapor-load coefficient. For flood ratios between 0.55 and 0.9, use:

$$C_{fL} = 1.75 - 1.4(\text{Flood ratio})$$



PARAMETER FOR $(HTU)_L$

PARAMETER FOR $(HTU)_V$

Figure 3-5. Packing Parameters for Determination of HTU[26]

NTU is shown as a function of stripping factor for a range of removal efficiencies in Figure 3-6.[3] An important observation made from this Figure is that for compounds with high Henry's Law constant and consequently high stripping factor (>50), the number of transfer units is nearly independent of the magnitude of Henry's Law constant for any given removal efficiency. It seems, therefore, that for compounds with high Henry's Law constant, NTU is proportional to removal efficiency alone. Note that the ratio of vapor to liquid flow rate does not have a significant impact on the stripping factor when Henry's Law constant is large.

3.2.4. Pressure Drop

Packed towers are usually designed to operate with a gas pressure drop well below flooding conditions. Most stripping towers are designed for gas pressure drops of 0.25 to 0.5 inches water per foot of packed height. In towers packed with rings or saddles, flooding typically occurs at a pressure drop between 2 and 3 inches per foot of height.[27]

The total gas pressure drop in a packed tower can be due to the following resistances:

1. Pressure drop due to the irrigated packing height. The generalized pressure-drop correlation shown in Figure 3-4 can be used to make estimates if manufacturer's data is not available.

2. Pressure drop for the dry packing height used above the liquid inlet as an entrainment separator. The pressure drop of a single fluid flowing through a bed of packed solids, when it alone fills the voids in the bed, is correlated by the Ergun equation. Approximately, for packed bed applications, this can be written:

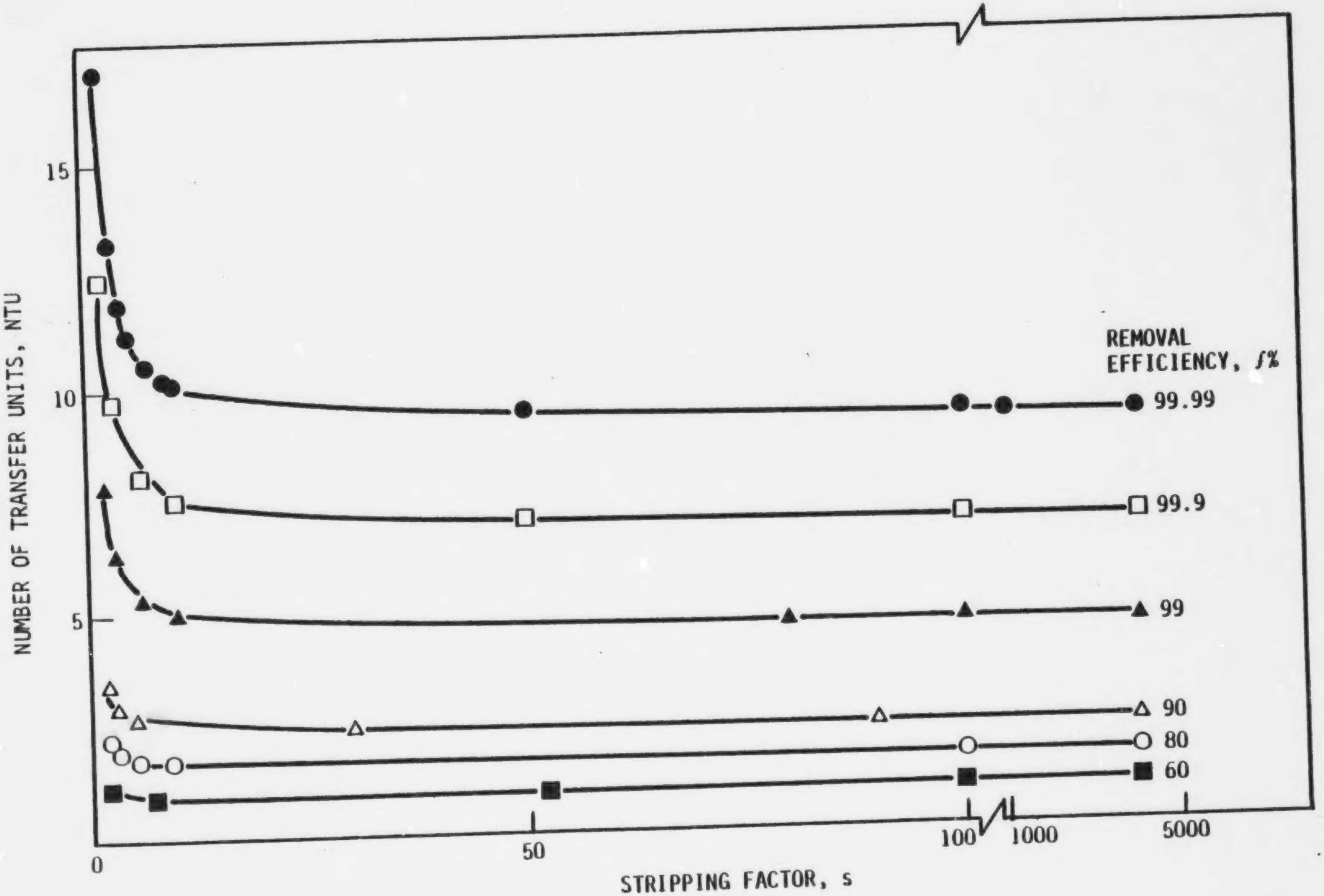


Figure 3-6. Dependency of NTU on Stripping Factor and Removal Efficiency[3]

$$\Delta P = 1.4\text{x}10^{-9}\ C_f(G')^2 Z/\rho_G \quad lb_f/ft^2 \qquad (3.35)$$

where G' is in lb/(hr ft^2). C_f is obtained from Table 3-4.

3. Pressure drop for the packing supports and liquid distributor, and for inlet expansion and outlet contraction losses for the gas. These may be neglected in comparison to the pressure drop across the irrigated packing.

3.3 ANCILLARY EQUIPMENT

3.3.1. Heat Exchangers

The design parameter of interest is the total heat transfer surface area which is the primary determinant of cost. In general, the surface area required, **A**, increases with increasing quantity of heat to be transferred and decreasing temperature difference, Δ**T**. The basic design equation is:

$$A = Q/(U\ \Delta T_m) \qquad (3.36)$$

This equation is based on the tube outside surface area and it provides a relationship between the three parameters that determine the heat transfer area required:

1. The total amount of heat to be transferred Q is fixed by process demand and is calculated from sensible and latent heat requirements.

2. **U** is the overall heat transfer coefficient. Table 3-6 provides values of **U** for a variety of applications encountered in stripping, and may be used for preliminary estimating purposes.[28] These values represent typical values found in the chemical industry and therefore a range is given. Values in the lower end of the range should be used for conservative estimates.

3. The temperature difference ΔT between the two streams. This normally varies throughout the length of exchanger, so the log mean temperature difference is used:

$$\Delta T_m = (\Delta T_2 - \Delta T_1)/\ln(\Delta T_2/(\Delta T_1) \qquad (3.37)$$

The temperature differences used in calculating the log mean are those at the inlet and outlet of the exchanger. Because the log mean temperature difference is based on ideal cocurrent or countercurrent flow, it is necessary to apply a correction factor, to allow for flow paths in actual exchangers. The type of exchanger used is normally selected so that the correction factor is greater than about 0.8. The reader is referred to Reference 28 for details on exchanger geometries and associated correction factors.

3.3.2. Storage Vessels

Storage vessels can be designed from the following factors:

1. The volumetric flow rate determined from material and energy balances.

2. The residence time of the fluid in the vessel. This can be the minimum time for which the fluid has to be stored, for example feed wastewater storage in the case of a stripping unit shutdown for emergency repairs.

3. A fixed over capacity factor, usually 20 percent.

The volume of a storage vessel can be calculated as follows:

$$\text{Volume} = 1.2\ V\ t \qquad \text{(gallons)} \qquad (3.38)$$

where V = volumetric flow rate, gallons/hr
t = residence time, hours.

TABLE 3-6

TYPICAL MEAN VALUES OF THE OVERALL HEAT-TRANSFER COEFFICIENT,[28] U

(Btu)/(^{o}F.ft^2.hr)

Shell side	Tube side	Design U
Liquid-liquid Exchangers		
Water	Water	200-250
Demineralized water	Water	300-500
Condensing vapor-liquid		
Steam	Feed water	400-1000
Gas-liquid Exchangers		
Air, compressed	Water or brine	40-80
Air, atmospheric	Water or brine	10-50
Water or brine	Air, compressed	20-40
Water or brine	Air, atmospheric	5-20

SECTION 4

COST ANALYSIS

Presented in this chapter are data and methods to allow estimation of capital costs and annualized costs for the various components of air and steam stripping systems. These costs can be used to determine the economic viability of stripping, evaluate alternative waste treatment processes, and for budget planning.

The purpose of the estimate determines the accuracy required, and in turn how much time and money is spent on it. Estimates have been classified and their accuracy standardized, as listed in Table 4-1, by the American Association of Cost Engineers. The potential limitations of these cost estimates must be recognized. Capital cost estimates of equipment that may be derived from the methods in this report are considered to be of study grade or better.

TABLE 4-1. AACE CLASSIFICATION OF CAPITAL COST ESTIMATES

Type of Estimate	Accuracy, ±%
Order of magnitude	40
Study	25
Preliminary or budget	12
Definitive or project control	6
Detailed or firm	3

Information on process equipment costs has been published in various engineering books, journals and several EPA reports. The cost data and methods presented in this report have been derived from these sources, and not from vendor quotes or case histories.

4.1 CAPITAL COSTS

Capital costs are the costs of the equipment used. They are discussed in terms of purchased cost, delivered cost and installed cost. Pikulik and Dias[29] suggest that the delivered cost of equipment may be approximated by increasing the purchased cost of equipment, f.o.b. manufacturer's shop, by 3 percent. An installation factor, usually different for each type of equipment, can be utilized to determine the installed capital cost. Installation factors are usually based on the purchased equipment cost.

The most accurate method of determining equipment costs is to obtain firm bids from manufacturers or suppliers. Often manufacturers can supply quick estimates which will fall close to the bid price. Second best in reliability are costs from the file of past purchase orders. When used for costing new equipment, purchase-order costs must be corrected to the current cost indexes.

It is often necessary to estimate the cost of a piece of equipment when no cost data are available for the particular size required. The logarithmic relationship known as the six-tenths-factor rule can be used if the new piece of equipment is similar to one of another capacity for which cost data are available.[14]

$$\text{Cost}_1 = \text{Cost}_o \; (\text{Size}_1/\text{Size}_o)^{0.6} \qquad (4.1)$$

However, the application of this rule of thumb for most purchased equipment is an oversimplification since actual values of the cost capacity factor vary from less than 0.2 to greater than 1.0.

Typical exponents are presented in this report for major stripping components. In general, the cost-capacity concept should not be used beyond a tenfold range of capacity. Care must be taken to make certain the two pieces of equipment are similar with regard to type of construction, material of construction, and temperature and pressure operating range.

Purchased equipment costs can often be estimated on the basis of weight. The fact that a wide variety of types of equipment have about the same cost per unit weight is quite useful, particularly when other cost data are not available. For fabricated vessels, such as stripping towers, fairly accurate costs can be derived on the basis of their weights. The procedures for translating the process design data from the previous Section into vessel weight equivalents are presented here.

The capital cost of air/steam stripping systems can be grouped into costs for the following major components, each of which is discussed separately:

1. Mass transfer equipment (tray and packed towers)

2. Heat transfer equipment (heat exchangers, condensers and reboilers)

3. Fluid transfer and handling equipment (pumps, compressors and tanks)

4. Installation materials including foundation, structural, instrumentation and controls, paint, insulation, electrical and piping, as well as labor - estimated by means of factors related to the purchased equipment costs.

All the capital cost data presented here are based on May 1983 cost indexes unless otherwise stated.

4.1.1. Mass Transfer Equipment

The purchased cost for tray and packed towers can be divided into the following components:

1. Shell cost, including heads, skirts, manholes and nozzles,

2. Cost for internals, including trays and accessories, packing, supports, and plates, and

3. Cost for auxiliaries, such as platforms, ladders, handrails, and insulation.

The published methods for determining these costs are largely graphical. However, with the advent of computer aided design and cost analysis, the use of correlations is becoming more convenient.

Hall et al[30] present purchased costs of tray type and packed towers in a series of graphs, as a function of diameter, height and specific operating parameters. The probable accuracies of these costs are in the range of ±10 to 15 percent.

A method for determining installed costs of plate type distillation towers, that includes items such as bulk material and field labor, overhead, engineering and contingencies, is presented by Miller and Kapella[31]. The base cost of the column is presented graphically as a function of the product of diameter, height and thickness. These costs represent Union Carbide's actual purchasing experience and can provide a reasonably accurate estimate (within 25 percent) for a completely installed tower.

The costs for essential components of a stripping tower provided in this report are in terms of correlations that were developed by Corripio et al[32] through the use of computer-aided statistical analysis of cost data for 200 distillation towers and 200 adsorption towers. In these correlations, tower cost represents the sum of the costs of the shell, platforms and ladders, and internals, either trays or packing. The costs of the shell are calculated from a knowledge of the shell weight.

The shell cost data on which the correlations are based include the cost of the skirt and a standard number of nozzles and manholes. These are functions of tower diameter, length and pressure rating. The shell weight is calculated (assuming 2:1 elliptic heads and ignoring the nozzles, manholes and skirt) from tower diameter, tangent-to-tangent length and design pressure (external or internal) by the procedure used for designing pressure vessels.[33] This procedure, as outlined below, provides a set of analytical equations and takes into account wind load effects and allows for different shell thickness at the bottom and top of the tower.

Shell weight

Shell weight depends on material density, diameter, tangent-to-tangent length and wall thickness. Thickness is a function of design pressure, diameter, length, and either the tensile strength or the modulus of elasticity of the material of construction (depending on whether the design pressure is positive or negative). For towers, the thickness requirement to withstand wind resistance must also be taken into consideration. The equation to calculate shell weight is:

$$W_s = D_i(L + 0.81D_i)\, T_s \rho_s \qquad (4.2)$$

where W_s = Shell weight, lb
D_i = Inside tower diameter, ft
L = Tangent-to-tangent length, ft
T_s = Tower wall thickness, ft, and
ρ_s = Density of shell material, lb/ft^3

Wall thickness for internal pressure. If the pressure in the tower is positive, the stress on the longitudinal seam governs. The maximum allowable stress S, which is a property of the specified material of construction, is then used in the following equation to calculate the wall thickness:

$$T_p = P_g(D_i/2)/(SE - 0.6\ P_g) \qquad (4.3)$$

where T_p = Thickness to withstand design pressure, ft
P_g = Gage pressure, psig
S = Maximum allowable stress, psi, and
E = Joint efficiency

Criteria for the selection of materials of construction are discussed later in this section. The wall thickness is one criteria in selecting the grade or quality of material. For example, in the case of carbon steel, the lower grade SA-285 C can be used for wall thickness up to 2 inches, and the higher grade SA-515-65 can be used for wall thickness greater than 2 inches.

The welding joint efficiency, E, is determined from tests in accordance with ASME codes. For carbon steel up to 1.25 inches in thickness, the weld need only be 10 percent spot-checked by X-rays, and a joint efficiency of 0.85 must be used in the formula. For any other thickness and for other materials of construction, the weld must be 100 percent X-ray tested and the joint efficiency is 1.0.

Wall thickness for external pressure. An iterative graphical procedure to calculate thickness under vacuum conditions is outlined in several sources. A method suitable for computer calculation is given here. The thickness required to withstand external pressure, T_p, is given by:

$$T_p = T_e + (T_e)_c \qquad (4.4)$$

T_e must be high enough for the collapsing pressure, P_c, given by Eq. (4.5) to be five times the difference between the external (atmospheric) pressure and the design (vacuum) pressure in the tower.

$$P_c = [2.6(T_e/D_o)^{2.5}E_M]/[(L/D_o - 0.45(T_e/D_o)^{0.5}] \quad (4.5)$$

where D_o = Outside tower diameter
E_M = Modulus of elasticity

The correction factor $(T_e)_c$ is obtained from the following equation, with T_e, L and D_i all in inches:

$$(T_e)_c = L(0.18\ D_i - 2.17)\text{x}10^{-5} - 0.19 \quad (4.6)$$

Thickness for wind load. The required wall thickness of a tower normally varies from top to bottom. The thickness at the top is that required to withstand either the internal or external pressure. The thickness at the bottom is additionally required to withstand the wind load, which is assumed to be sufficient also for any earthquake load. Assuming that the wind acts with a uniform intensity over the entire length of the column and that the drag coefficient for the wind resistance is 1.0 (drag past a cylinder in turbulent flow), the thickness necessary to withstand the wind load is calculated from:

$$T_w = \rho_a V^2(D_o + Z)L^2/(S\pi D_o^2) \quad (4.7)$$

After substituting common values for some variables such as:

ρ_a = Air density, 0.075 lb/ft^3 at 70°F and 1 atm,
V = Wind velocity, 140 miles/hour, and
Z = Allowance for cage ladders, 1.5 ft.

we get the following simplified equation with T_w, D_o, and L in inches, and S in psi:

$$T_w = 0.22(D_o + 18)L^2/(SD_o^2) \qquad (4.8)$$

The thickness required to withstand the internal pressure when the girth seam governs is calculated by:

$$T_g = P_g(D_i/2)/(2SE + 0.4\ P_g) \qquad (4.9)$$

The total thickness at the bottom of a tower is then given by:

$$T_b = T_w + T_g \qquad (4.10)$$

Finally, the total shell thickness is taken to be the greater of the calculated top and bottom thicknesses, plus a corrosion allowance, T_c, which is specified by the user.

$$T_s = (\text{greater of } T_b,\ T_p) + T_c \qquad (4.11)$$

This final calculated thickness should be rounded up to the nearest standard plate thickness.

Shell cost

The shell cost or the base cost of the tower, which covers fabrication and prime painting in the shop, is f.o.b. manufacturer's plant. Correlations for base cost, in carbon steel, of the shell are given for both tray towers and packed towers in Table 4-2. To calculate shell cost for a material of construction other than carbon steel, use the material-of-construction factors, F_M, also given in Table 4-2, as a multiplier to the base cost in carbon steel.

To fabricate a shell of thickness varying from top to bottom, the additional labor cost required is significant only for towers having large length-to-diameter ratios, as these must be thicker at the bottom to withstand wind loading. This is evident in Eq. (4.13) for towers taller than 40 ft. The

TABLE 4-2

CORRELATIONS FOR COST* OF STRIPPING TOWERS[32]

Carbon steel shell, base cost C_B, $

$$C_B = 1.3\exp[6.3 + 0.18\ \ln W_s + 0.02\ \ln(W_s)^2] \qquad (4.12)$$

for $L_t \leq 40$ ft, and
$4250 \text{ lb} \leq W_s \leq 980{,}000 \text{ lb}$

$$C_B = 1.3\exp[6.8 + 0.14\ \ln W_s + 0.02(\ln W_s)^2 + 0.016(L/D_i)\ln(T_b/T_p)] \qquad (4.13)$$

for $L_t > 40$ ft
$9020 \text{ lb} \leq W_s \leq 2{,}470{,}000$

Final shell cost C_S, $

$$C_S = F_M\ C_B \qquad (4.14)$$

where

Material	Cost factor, F_M
Carbon steel	1.0
Stainless steel, 304	1.7
Stainless steel, 316	2.1
Carpenter 20CB-3	3.2
Nickel - 200	5.4
Monel - 400	3.6
Inconel - 600	3.9
Incoloy - 825	3.7
Titanium	7.7

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* Costs were escalated to May 1983 using the Chemical Engineering Fabricated Equipment Index of 327.1

requirement is smaller, however, for tall towers of higher design pressure because of the greater thickness at the top of such towers. The ratios of tower length to diameter and bottom to top thickness in the cost correlations accounts for this additional labor cost requirement. Eqs. (4.12) and (4.13) are each based on about 200 data points on the cost of distillation and adsorption towers covering a wide range of design variables and materials of construction. The standard deviation is 9.9 percent for Eq. (4.12) and 10.6 percent for Eq. (4.13).

Platforms and Ladders

Correlations for the base cost, C_L, for carbon steel platforms and ladders are given in Table 4-3. The standard deviation for Eq. (4.15) is 8.9 percent and that for Eq. (4.16) is 3.4 percent. The significant difference between the two standard deviations reflect the effect of correlating the discretely varying number (and cost) of platforms with a continuous variable, tower length. The effect of this error is much larger for the shorter towers than for the taller towers.

Cost of Internals

Cost of trays, C_T. Correlations for the base cost of trays were developed from cost data for Glitsch "Truss type" one-pass removable ballast trays. The base cost of valve trays in carbon steel are given as functions of tower diameter in Table 4-4. For the correlation of 14 trays of different diameter, the standard deviation is 1.3 percent.

Material-of-construction cost factors, F_{MT}, and tray-type factors, F_T, that must be applied to the base cost are given in Table 4-4. If a design calls for fewer than 20 trays, a number-of-trays factor, F_N, recommended by Enyedy[34] must be applied:

$$F_N = 2.25/(1.0414)^N \qquad (4.19)$$

TABLE 4-3 CORRELATIONS FOR COST OF PLATFORMS AND LADDERS[32]

$$C_L = 237\ D_i^{0.74}\ L^{0.71} \qquad (4.15)$$

for $27\ \text{ft} \le L \le 40\ \text{ft}$

$3\ \text{ft} \le D_i \le 21\ \text{ft}$

$$C_L = 198\ D_i^{0.6}\ L^{0.8} \qquad (4.16)$$

for $57\ \text{ft} \le L \le 170\ \text{ft}$

$3\ \text{ft} \le D_i \le 24\ \text{ft}$

TABLE 4-4. CORRELATIONS FOR COST OF TOWER TRAYS[32]

Carbon steel valve tray base cost C_{BT}, $

$$C_{BT} = 362\ \exp(0.174\ D_o) \qquad (4.17)$$

for $2\ \text{ft} \le D_o \le 16\ \text{ft}$

Final cost $C_T = C_{BT}\ F_{MT}\ F_T$ (4.18)

Material	Cost Factor, F_{MT}
Carbon steel	1.0
Stainless steel, 304	$1.19 + 0.06\ D_i$
Stainless steel, 316	$1.40 + 0.07\ D_i$
Carpenter 20 CB-3	$1.53 + 0.08\ D_i$
Monel	$2.31 + 0.11\ D_i$

Tray Type	Cost Factor, F_T
Valve	1.0
Grid	0.8
Bubble cap	1.59
Sieve (with downcomer)	0.85

Cost of tower packings, C_P. Cost of tower packings are based on data from Pikulik and Diaz[29]. The costs are estimated from the required volume of packing and its cost per unit volume as listed in Table 4-5. The cost of a distributor plate in a packed tower can be assumed to be the same as that for one bubble cap tray.

TABLE 4-5. COST OF TOWER PACKING PER UNIT VOLUME[29]

Packing type	C_P, \$/ft^3
Ceramic Raschig rings or Intalox saddles, 1 in.	18.9
Ceramic Raschig rings or Intalox saddles, 2 in.	13.2
Metal Raschig or Pall rings, 1 in.	31.1
Metal Raschig or Pall rings, 2 in.	22.1

Total Purchased Cost

Total cost of tray towers is the sum of the costs for individual components as discussed in previous sections. If **N** is the number of trays required, then the total cost can be estimated using the following equation:

$$C = C_S + C_L + N\,C_T \tag{4.20}$$

Similarly, the total cost of a tower packed to height **Z** is:

$$C = C_S + C_L + (\pi D_i^2/4) Z\,C_P \tag{4.21}$$

Correlations for the cost of towers having two diameters have been published by Enyedy.[34]

4.1.2 Heat transfer equipment

The basic capital cost for shell-and-tube heat transfer equipment - which includes heat exchangers, condensers and reboilers - is a function of many design and cost parameters. The most important parameters are exchanger type (fixed tubesheet, U-tube, floating head and kettle-type), heat transfer surface area, diameter, lengths and gauge of the tubes, design pressure, material of construction, and the type and inlet temperature of the fluids. The fixed tubesheet design is low in cost and cleanable on the tubeside. It should not be used in applications involving severe thermal expansion stresses and severe fouling. The U-tube design is moderate in cost and eliminates the problem of thermal expansion. However, they impose difficulty in tubeside cleaning. The floating head design is the highest in cost, but both the tubes and the shell are easily cleaned. If low temperatures are required for condensation, refrigeration can add significantly to the total capital and operating costs of a heat transfer system.[40]

Costs of shell-and-tube heat exchangers are correlated against heat transfer area and are presented by Corripio *et.al.*[41] The base cost is for floating-head shell-and-tube exchangers made of carbon steel and designed for 100 psig pressure. Cost factors for design type, design pressure and material-of-construction are provided separately and must be used to adjust the base cost. Table 4-6 lists correlations, factors and limits for their use.

The cost estimation procedure that takes into account shell diameter, number, length and gauge of tubes, types of heads and other construction details can be more accurate and is discussed in several References.[14,40] The accuracy of the correlation of cost vs. area provided here is adequate for study cost estimates.

TABLE 4-6

CORRELATIONS FOR COSTS OF HEAT TRANSFER EQUIPMENT[41]

Base cost for floating-head, carbon steel, 100 psig design pressure*:

$$C_B = 1.3\exp[8.55 - 0.31(\ln A) + 0.07(\ln A)^2] \qquad (4.22)$$

Final cost:

$$C_E = C_B\ F_T\ F_P\ F_M \qquad (4.23)$$

Type	Factor F_T
Fixed tube-sheet	$\exp[-1.12 + 0.091 \ln A]$
U-tube	$\exp[-0.98 + 0.083 \ln A]$
Kettle reboiler	1.35

Pressure	Factor F_P
100 to 300 psig	$0.78 + 0.05 \ln A$
300 to 600 psig	$1.03 + 0.07 \ln A$
600 to 900 psig	$1.14 + 0.12 \ln A$

Material	Factor F_M
Carbon steel	1.0
Stainless Steel, 316	$0.86 + 0.23 \ln A$
Stainless Steel, 304	$0.82 + 0.16 \ln A$
Stainless Steel, 347	$0.61 + 0.22 \ln A$
Nickel-200	$1.51 + 0.61 \ln A$
Monel-400	$1.30 + 0.43 \ln A$
Inconel-600	$1.20 + 0.51 \ln A$
Incoloy-825	$1.19 + 0.50 \ln A$
Titanium	$1.54 + 0.43 \ln A$
Hastelloy	$0.15 + 1.52 \ln A$

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* A in ft^2; $150 \leq A \leq 12{,}000$

4.1.3 Centrifugal Pumps

Pumps used for transferring aqueous dilute solutions are commonly centrifugal-type and the easiest way to estimate their cost is by using a cost-data manual which all pump manufacturers provide to their customers. These manuals are detailed and give reliable estimates. For designs not covered in the manual, the usual procedure is to send detailed specifications to several pump manufacturers for their bids.

Correlating the cost of centrifugal pumps against size or capacity is difficult because a pump of a given capacity can serve in a variety of combinations of flowrate (capacity) and developed head. This difficulty is circumvented by correlating pump cost against the maximum value of the size parameter S, defined below, that could be handled by a pump of a particular cost.

$$S = Q\ H^{0.5}$$

Here Q is design capacity in gpm, and
H is the required head in ft-lb/lb.

The base cost is for a one-stage, 3550 rpm, vertically-split-case (VSC) pump of cast iron. Correlations for pump base cost and design-type factors, and cost factors for material-of-construction are given in Table 4-7.[42] The cost of a pump calculated from these correlations includes the base plate and driver coupling but not the driver. The capacity, head and horsepower limits for each type of pump are noted in Table 4-8.

TABLE 4-7. CORRELATIONS FOR COST* OF CENTRIFUGAL PUMPS[42]

Base Cost C_B for single-stage, 3550 rpm, cast iron, VSC:

$$C_B = 1.53\exp[8.4 - 0.6(\ln S) + 0.05(\ln S)^2] \qquad (4.24)$$

Total cost:

$$C_P = C_B\, F_T\, F_M \qquad (4.25)$$

Type	Factor F_T
1-stage 3550 rpm, VSC	1.0
1-stage 1750 rpm, VSC	$\exp[5.10 - 1.22 \ln S + 0.077(\ln S)^2]$
1-stage 3550 rpm, HSC	$\exp[0.06 + 0.27 \ln S - 0.025(\ln S)^2]$
1-stage 1750 rpm, HSC	$\exp[2.03 - 0.24 \ln S + 0.010(\ln S)^2]$
2-stage 3550 rpm, HSC	$\exp[13.73 - 2.83 \ln S + 0.154(\ln S)^2]$
Multistage 3550 rpm, HSC	$\exp[9.88 - 1.62 \ln S + 0.083(\ln S)^2]$

Material	Factor F_M[26]
Carbon steel	1.0
Cast steel	1.35
304 or 316 fittings	1.15
Stainless steel, 304 or 316	2.00
Cast Gould's Alloy No. 20	2.00
Nickel	3.50
Monel	3.30
ISO B	4.95
ISO C	4.60
Titanium	9.70
Hastelloy C	2.95
Ductile iron	1.15
Bronze	1.90

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*Costs were escalated to May 1983 using Chemical Engineers's Pumps and Compressors Index of 413.4.

TABLE 4-8

CAPACITY, HEAD AND HORSEPOWER LIMITS FOR CENTRIFUGAL PUMPS[42]

Type	Flow gpm	Head ft-lb_f/lb	Motor hp (max)
1-stage 3550 rpm, VS	50 to 900	50 to 400	75
1-stage 1750 rpm, VSC	50 to 3,500	50 to 200	200
1-stage 3550 rpm, HSC	100 to 1,500	100 to 450	150
1-stage 1750 rpm, HSC	250 to 5,000	50 to 500	250
1-stage 3550 rpm, HSC	50 to 1,100	300 to 1,100	250
Multistage 3550 rpm, HSC	100 to 1,500	650 to 3,200	1,450

4.1.4 Electric motors

To determine the cost of a driver for a pump, the required brake horsepower is determined using the following formula:

$$P_B = Q\,H/(33{,}000\ E_p)$$

where,
ρ = fluid density, lb/gal
Q = liquid flow, gpm
H = developed head, ft-lb_f/lb

Pump efficiency E_p may be calculated from:

$$E_p = -0.316 + 0.24(\ln Q) - 0.012(\ln Q)^2$$
$$19 \leq Q,\ \text{gpm} \leq 5000$$

The cost correlations and coefficients for motors are given in Table 4-9 and are for three motor types and three speeds. Because electric motors come in discrete sizes, the motor size to use with Table 4-9 is the available motor horsepower size that is just equal to or bigger than the required brake horsepower.

TABLE 4-9 CORRELATIONS FOR COST* OF ELECTRIC MOTORS[42]

$$C_m = 1.38 \exp[a_1 + a_2(\ln P) + a_3(\ln P)^2] \qquad (4.26)$$

P = nominal power of motor, hp ($\geq$ brake horsepower P_B)
60 Hz, standard voltage motor & insulation

	a_1	a_2	a_3	hp limit
Open, drip-proof				
3,600 rpm	4.83	0.097	0.11	1-7.5
	4.15	0.535	0.053	7.5-250
	4.24	1.033	-0.036	250-700
1,800 rpm	4.71	-0.015	0.229	1-7.5
	4.52	0.472	0.048	7.5-250
	7.40	-0.065	0.054	250-600
1,200 rpm	4.93	0.301	0.126	1-7.5
	5.10	0.359	0.061	7.5-250
	4.62	0.885	-0.022	250-500
Totally enclosed, fan-cooled				
3,600 rpm	5.11	0.033	0.154	1-7.5
	3.85	0.833	0.024	7.5-250
	5.32	1.085	-0.057	250-400
1,800 rpm	4.97	-0.009	0.226	1-7.5
	4.53	0.571	0.046	7.5-250
1,200 rpm	5.15	0.289	0.144	1-7.5
	5.39	0.310	0.074	7.5-350
Explosion-proof				
3,600 rpm	5.39	-0.003	6.155	1-7.5
	4.44	0.608	0.052	7.5-200
1,800 rpm	5.29	0	0.200	1-7.5
	4.82	0.511	0.053	7.5-250
1,200 rpm	5.42	0.312	0.106	1-7.5
	5.57	0.313	0.072	7.5-200

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*Costs were escalated to May 1983 using Chemical Engineer's Electrical Equipment Index of 242.4

4.1.5 Air compressors

Air stripping requires a compressed air supply. Compressors having high capacity and low discharge pressures are required. Centrifugal air compressors will adequately do the job. Their cost, C_c, can be determined as a function of capacity using the correlation fitted to data from Peters and Timmerhaus.[13]

$$C_c = 1.53 \times \exp[0.31(\ln Q) + 8.52] \qquad (4.27)$$

Q = capacity, ft^3/min; $50 \le Q \le 1000$

Typical speeds and horsepower for various capacities are:

Q	rpm	HP
143	230	50
275	275	75
325	300	100

The costs were escalated to May 1983 using Chemical Engineering's Pumps and Compressors Index of 413.4.

4.1.6 Storage vessels

Preliminary-grade estimates of storage vessels can be made using the correlations and factors for material-of-construction given in Table 4-10. Total vessel or tank volume is calculated from the knowledge of tank residence time, a fixed overcapacity factor of 20 percent, and volumetric flow rate. The last is determined from material and energy balance calculations. Omitted in the cost estimating procedure are the number and sizes of manholes and nozzles and other design details.

TABLE 4-10

CORRELATION FOR COSTS OF STORAGE VESSELS[41]

Base cost for shop-fabricated carbon steel tanks:

$$C_B = 1.3 \exp[2.33 + 1.37(\ln V) - 0.063(\ln V)^2] \qquad (4.28)$$

$$1{,}300 \leq V \text{ gallons} \leq 21{,}000$$

Base cost for field-erected carbon steel tanks:

$$C_B = 1.3 \exp[11.4 - 0.610(\ln V) + 0.045(\ln V)^2] \qquad (4.29)$$

$$21{,}000 \leq V \text{ gallons} \leq 11{,}000{,}000$$

Final Cost:

$$C_V = C_B F_M \qquad (4.30)$$

Material	Factor F_M
Carbon steel	1.0
Stainless steel 316	2.7
Stainless steel 304	2.4
Stainless steel 347	3.0
Nickel	3.5
Monel	3.3
Inconel	3.8
Zirconium	11.0
Titanium	11.0
Brick-and-rubber or brick-and polyester lined steel	2.75
Rubber- or lead-lined steel	1.90
Polyester, fiberglass-reinforced	0.32
Aluminum	2.70
Copper	2.30
Concrete	0.55

Judgment on optimum allowance for corrosion and on the use of the most suitable construction materials needs experience or vendor assistance. Guidance can be obtained by reference to a materials handbook.[47]

It is customary to have spare pumps and motors available. In larger plants, standby pumps may be included in the process flow sheet on by-pass lines. For a stripper facility, consider having back-ups for all pumps and motors as well as the air compressor. These must be included in the capital cost estimate.

4.1.7. Installation Costs

The installation of equipment involves costs for labor, foundations, supports, platforms, construction expenses, and other factors directly related to the erection of purchased equipment. The following list of additional items/materials must be included in estimating the total cost of installation materials:

- Equipment insulation
- Instrumentation and controls
- Piping and insulation
- Contractors fees
- Contingencies
- Service facilities
- Engineering, Supervision and Construction
- Electrical installations
- Buildings
- Yard improvement
- Land
- Start-up

Capital costs of equipment, system or plant, that include costs for all of the above items are often referred to as the total investment cost. From experience one can generate factors, as a percent of purchased equipment cost (or some other convenient base cost), for each of the above categories to estimate the investment cost or "turnkey" cost for a system. We have not found factors for air and steam stripping systems in the literature.

There are several methods available to estimate the total investment cost of a system when installation costs are not available. These include the methods of Lang,[35] Hand,[36] Guthrie[37] and Viola.[38] In Lang's method the total investment cost of a stripping system is estimated by multiplying the total delivered cost of equipment by a factor of 4.74. As discussed above, the delivered cost of equipment may be approximated by increasing the purchased cost, f.o.b. manufacturer's shop, by 3 percent.

In Hand's method,[36] different installation factors are used for each type of equipment. Some typical factors[39] based on the purchased equipment cost are:

Equipment	Factor
Towers	4
Heat exchangers	3.5
Compressors	2.5

The purchased equipment cost must be multiplied by these factors to arrive at total installed costs, and the sum of these products represents the estimated inside-battery-limit cost of the installed plant. While these methods are convenient to use they provide, at best, study estimates.

4.2 Operating Costs

Determination of capital costs is only one part of a complete cost estimate. Another equally important part is the estimation of costs for operating the system and maintaining it to run efficiently. Operating costs include both fixed and variable components.

Total operating costs are commonly calculated on one of three bases: daily basis, unit-of-product basis, or annual basis. The annual cost basis is probably the best choice because:

1. The effect of seasonal variation is smoothed out, and
2. The plant stream factor is included.

Fixed operating costs include labor, supervision, overhead, laboratory labor, maintenance, services, insurance and taxes, service water, and the cost of capital. Variable components include utilities - electricity, steam, cooling water, and compressed air.

4.2.1 Fixed Costs

The bases and factors for fixed operating costs are given in Table 4-11 obtained from the EPA cost model. Cost factors in the original are for July 1977; values adjusted with appropriate cost indexes to May, 1983 are also shown in the Table.

4.2.2 Variable Costs

Power Requirements

These include power for the feed, bottoms, reflux, and overhead liquid pumps, and for the air compressor. The pump horsepower is estimated using:

$$HP = 1.21 \times 10^{-6} \times Q \times H \qquad (4.28)$$

where Q is the flow rate, lb/hr, and
H is the head, ft.

A 20 percent factor for flow variation and a 50 percent pump efficiency is assumed. The following heads may be used:

Feed-pump head	50 + L ft
Bottoms-pump head	50 + L/2 ft
Overhead and reflux pumps	20 ft

L is the total tower height in feet.

TABLE 4-11

FIXED OPERATING AND MAINTENANCE COST BASIS AND UNIT COST FACTORS FOR STRIPPING[8]

Element	Cost Basis Equivalent Unit Quantity	Base Unit Cost July 1977	 May 1983
Labor*+	0.25Weeks(6hr/day)	\$ 9.80/hr	\$14.70
Supervision*	10% Labor (0.60 hr/day)	\$11.76/hr	\$23.25
Overhead*	75% Labor Cost	NA	NA
Laboratory**	0.20 Shifts (1.6 hr/day)	\$10.70/hr	\$16.00
Maintenance	4.13% Capital	NA	NA
Services	0.40% Capital	NA	NA
Insurance and Taxes	2.50% Capital	NA	NA
Service Water	10^3 gal/day	\$0.50/10^3 gal	\$1.00/10^3 gal
Amortization	16% Capital (estimated)	NA	NA

NA - not applicable

* Labor may vary from 0.7 to 1.2 times the standard amount indicated depending on the overall scale of the plant. Labor, Supervision, and Overhead may also be adjusted for the scale of the plant.

+ One week = 7 days = 168 hours = 4.2 shifts

** One shift = 40 hours

Note: The May, 1983 data were estimated using Chemical Engineering's indexes for construction labor.

The horsepower for the compressor can be calculated from the equation for an ideal gas undergoing isothermal compression:[14]

$$HP = 3.03x10^{-5} p_1 \ln(p_2/p_1) \qquad (4.29)$$

where, Q = air flow rate in ft^3/m

p_1, p_2 = intake and delivery pressure, lb_f/ft^2

Equations for isentropic compression should be used for high compression ratios.

The total annual power cost is given by:

$$C_p = 6535\ S_F\ P\ (\$/kWhr) \qquad (4.31)$$

where S_F is the stream factor and P is the sum of the horsepower requirements for all the pumps and compressors.

Utilities

The cooling water requirement is approximately equivalent to the latent heat of the steam and organics condensed in the condenser. Assuming a 20°F temperature rise and a 20 percent factor for flow variations, the cooling water requirement is:

$$CW = 63\ S_F\ (\text{condenser duty, Btu/hr})\ \text{gals/year} \qquad (4.32)$$

The steam rate is the design stripping medium rate plus any requirements for heating the feed and compensating for heat losses in the tower.

Costs for steam and cooling water are normally obtained from plant records. If data are not available, steam costs can be based on an energy cost of $\$3/10^6$ Btu, and a cooling water cost based on 3 to 5 cents/1000 gallons water circulated.

SECTION 5

SUMMARY OF DESIGN PROCEDURES

A step by step approach suitable for handling most design situations is given here to assist the reader through the numerous correlations and situations covered in the previous sections. An important part of obtaining a reliable design is having accurate physical and chemical property data available. Henry's Law constants and activity coefficients for the organics on the EPA's toxic pollutant list are included in the Appendix. These should be checked and updated with recent data as it becomes available. Lists of densities, viscosities, specific heats, latent heats and surface tensions for a range of species can be found in many handbooks.[28,43,44] Diffusivities are less easily found, but may be estimated using correlations and data given in Reference 17.

5.1 PROCESS DESIGN

1. Basic Design Data

The basic design data required includes the flow rate and temperature and pressure of the stream to be stripped. Temperature and pressure constraints due to chemicals in the stream or imposed by the process should be noted. A detailed stream composition is useful, particularly with respect to suspended matter and reactive or corrosive substances. Treatments to be provided upstream of the stripper should be identified.

Henry's Law constants for all the components should be obtained, preferably as a function of temperature as given in the Appendix. In the case of multicomponent streams, select a key component on which to base the design. Molecular weights will be required to convert the concentrations to mole fractions. Other physical property data that will be required are listed later.

2. Preliminary Tower Height Calculations

The design options available can be relatively quickly ascertained by doing a few equilibrium stage calculations using the simplified equations in Table 3-2. Determine the number of stages when stripping with steam at from 10 to 30 mole percent of the feed, and with air at from one-half to twice or more the feed rate (molar basis). Examine the effect of various amounts of reflux in the case of steam.

3. Selection of Stripping Medium

The preliminary designs will give an indication of the effectiveness of stripping and should also suggest whether air stripping is satisfactory or whether steam should be used. Steam will usually be less expensive if it means an 8-fold decrease in the amount of stripping medium used.[3] Steam may also be dictated for process reasons related to recovery of the organics and/or disposal of the overhead vapors. If the choice is not obvious, complete cost estimates for both steam and air, and select the less expensive system.

Live and reboil steam. In the case of steam, a choice must be made between live steam or the use of a reboiler. A reboiler adds to the capital costs, but the steam is recovered as a clean condensate and is recycled. An equivalent amount of vapor is driven off the feed in the overhead vapors, so the bottoms rate is less than the feed rate. Live steam does not significantly alter the feed stream rate. The live steam must be continually made up from the boiler feedwater treatment plant. In some situations, it may be possible to use a low purity steam raised from wastewater.

Condenser and reflux. The incorporation of a condenser in the system may also influence the decision between reboil and live steam. A condenser will be used wherever recovery of the organics is required. A partial condenser may be used to concentrate the overhead in systems that incinerate stripped vapors. The amount of reflux is determined by the quality of overhead required. When a decanter is used, all the aqueous phase may be returned to the tower as reflux.

4. Selection of Tower Type

Select between a packed or tray tower on the basis of the information given in Tables 2-1 and 2-2. This information together with the data in Table 3-4 and the discussion in Section 3.2.1 will assist in the selection of the type of packing or tray. Availability and costs should be ascertained by consulting the manufacturers.

5. Detailed Process Design

Sufficient background information is now available to carry out the process design. Tower capital costs decrease with increasing stripping medium, but operating costs increase. An optimum design having minimum overall costs can usually be found. It is therefore recommended that the designs be repeated for a range of stripping medium rates.

Tower height. For steam, use the equations in Table 3-2 for tray towers, and Table 3-5 together with Eqs. (3.31), (3.32) & (3.34) for packed towers. In the case of (adiabatic) air stripping, the number of equilibrium stages can be obtained using the BASIC program given in the Appendix. Because a similar program is not provided for packed towers, the isothermal equations may be used, but with some loss of accuracy.

The equation for the determination of HTU incorporates factors containing the surface tension, viscosity and density of the liquid relative to water. Density data is usually readily available, but surface tension and viscosity data may be less easily found. Although these latter variables can vary significantly with even small concentrations of organics, the HTU is not strongly dependent on them (they are raised to low powers). Neglecting these factors normally will not introduce large errors, especially when designing systems in which the organic is stripped down to trace amounts. The diffusion coefficient of the organic through the liquid is always required, and may be estimated using published correlations.[17]

Note that the determination of HTU and tower height for packed towers is an iterative procedure. The use of a programmable calculator or microcomputer is recommended.

Tower diameter. Liquid and vapor densities are required and, in addition, the liquid surface tension for tray towers, or the liquid viscosity for packed towers, should be known.

For tray columns, use Eqs. (3.26) to (3.28) and the parameters in Table 3-3. Manufacturer supplied data should be used if available. The tower diameter also sets the tray spacing (See Table 3-3), which, when multiplied by the number of stages, is the height required for the tray section of the tower. Additional height may be required for a reboiler, and an extra static head or 'skirt' height (typically about 15 ft) may be added to prevent cavitation in the bottoms pump.

For packed towers, the tower diameter is determined from Figure 3-4 as described in Section 3.2.2. If packing parameters are not available from the manufacturers, use the data in Table 3-4.

Stage Efficiency. For tray towers use the equations summarized in Table A-1, together with Figure A-1 in the Appendix. Extensive physical property data are again required. The actual number of trays required is given by the number of theoretical stages divided by E_o, the overall tower efficiency.

Pressure drop. The pressure drop is required for costing the pumps and air fan. Refer to Section 3.1.4 and 3.2.4 for tray and packed towers, respectively.

6. Ancillary Equipment

Obtain the surface area for all condensers, heat exchangers, reboilers, and storage vessels as outlined in Section 3.3. Refer to Figure 3-1 for an indication of the equipment required for a typical stripping plant.

5.2 COST ESTIMATION

The correlations given in Section 4 are provided as a guide to cost estimating. Wherever possible, costs from previous projects or vendor quotes should be used. The correlations are readily programmable for use on calculators or small computers.

Finally, the capital cost, operating cost and overall annualized cost should be plotted against the stripping medium rate, and the optimum design selected.

SECTION 6

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APPENDIX A

DETERMINATION OF TRAY EFFICIENCY

TABLE A-1

SUMMARY OF AIChE PROCEDURE FOR PREDICTION OF TRAY EFFICIENCY

1. Predict a value for $(NTU)_V$, the number of phase transfer units from:

$$(NTU)_V = 1.51/(Sc)^{0.5}$$

where Sc = dimensionless gas-phase Schmidt number
$= \mu_V/(\rho_G D_V)$
μ_V = gas-phase viscosity
ρ_G = gas-phase density
D_V = gas-phase diffusivity of the pollutant into water vapor

2. Compute the liquid residence time on a tray, t_L, seconds using:

$$t_L = 0.66 D$$

where D = tray diameter in ft

3. Predict a value for $(NTU)_L$, the number of liquid-phase transfer units, from:

$$(NTU)_L = 89.2 \ D \ (D_L)^{0.5}$$

where D_L = liquid-phase diffusivity of pollutant into water at nearly infinite dilution in water, ft^2/hr

4. Compute the overall gas-phase transfer units, NTU, from

$$1/NTU = 1/(NTU)_V + S/(NTU)_L$$

where S = stripping factor, KV/L

5. Predict the point efficiency:

$$E_P = 1 - \exp(-NTU)$$

6. Compute a value for effective diffusivity in the direction of liquid flow:

$$D_E = (10.034 + 1.026 \ v)^2 \qquad ft^2/hr$$

where v = vapor velocity through the active tray area, ft/s

7. Compute the Peclet number, Pe, from:

$$Pe = 3600 \ Z_1^2/(D_E \ t_L)$$

where Z_1 = distance traveled on the tray by the liquid, ft; may be taken as the distance between inlet and outlet weirs, = 0.71 x diameter.

8. Obtain the ratio E_M/E_P

$$E_M/E_P = \{1 - \exp[-(M + Pe)]\}/\{(M + Pe)[1 + (M + Pe)/M]\}$$
$$+ (\exp M - 1)/\{M[1 + M/(M + Pe)]\}$$

where $\quad M = (Pe/2)[(1 + 4\ S\ E_P/Pe)^{0.5} - 1]$

9. Obtain the fractional entrainment (mass fraction of liquid that is entrained), Ψ, from Figure A-1.

10. Correct the resulting tray efficiency for the effect of entrainment using Colburn's equation:

$$E_A = E_M/[1 + E_M\ \Psi/(1 - \Psi)]$$

11. The overall column efficiency can be related to the tray efficiency by the following relatioship:

$$E_o = \ln[1 + E_A(S - 1)]/\ln S$$

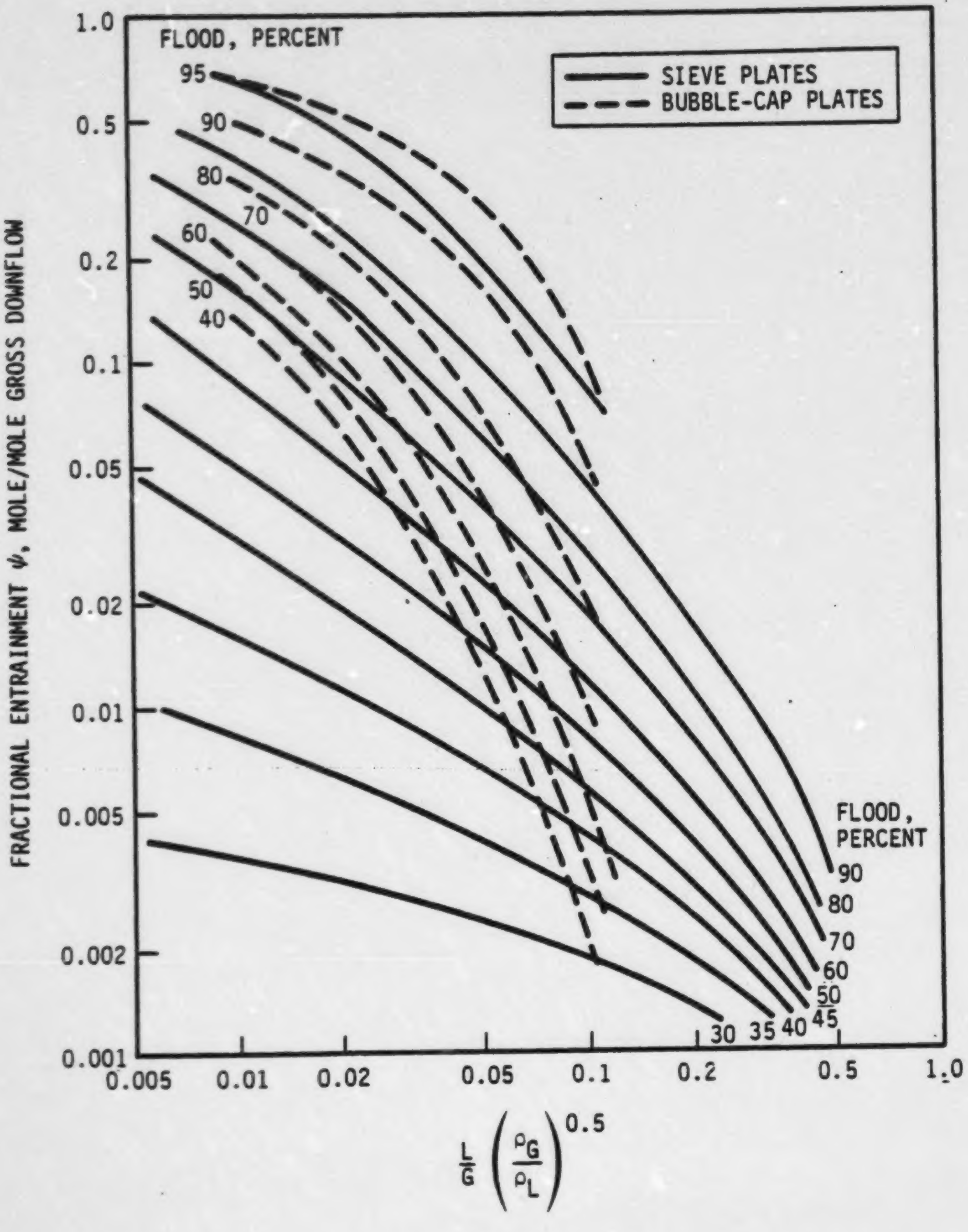

Figure A-1. Entrainment Correlation [10]

APPENDIX B

DESIGN EXAMPLE

A dilute aqueous solution of an organic is to be steam stripped from an initial mole fraction of 27×10^{-6} down to 0.6×10^{-9} Design the tower to handle a feed of 2100 moles/hr, and use a steam rate of 260 moles/hr (about 12.4 percent of the feed). Assume the column operates at a pressure of one atmosphere and 100°C, and that at these conditions the vapor-liquid equilibrium constant, K, has a value of 15.

1. TRAY TOWER DESIGN (Perforated trays)

Theoretical Stages

Stripping factor $S = KV/L = 15\times260/2100 = 1.86$

Removal efficiency $f = 1 - 0.6\times10^{-9}/27\times10^{-6} = 0.99998$

From Eq. (3.18) $N = 16.0$ for $R = 0$, and

$N = 8.73$ for $R = 99$

In the reflux case, it was assumed that 99 percent of the overhead would be recovered in a decanter and returned to the tower as aqueous reflux. Other values used were:

$t_B = 100^\circ C$ $t_R = 95^\circ C$

$\gamma_D = 1.0$ (nearly pure organic phase assumed)

$\gamma_S = 1131$ (values for specific compounds are tabulated in the Appendix and in Reference 20)

$C_p = 1.0$ cal/g.K)

$L = 542$ cal/g

which gives $R' = 99.9$ and $k = 0.0108$

Tower Diameter

For water at 100°C:

Liquiddensity= 60 lb/ft^3; Vapordensity=0.037 lb/ft^3
Surface tension = 59.2 dyne/cm
Liquid mass rate = 2100x18 = 37,800 lb/hr
Gas mass rate = 260x18 = 4680 lb/hr

To calculate the vapor velocity through the active tray area we use Eq. (3.28) with:

$A_h/A_a = 0.13$ (assumed)
Tray spacing, $t = 20$ inches
$a = 0.16$
$b = 0.10$
$c = 0.20$
$K_v = 0.24$

From Eq. (3.26), Maximum allowable velocity, $v_m = 12.2$ ft/s

The volumetric flow rate of the gas is 4680/(3600x0.037) or 35.1 ft^3/s

From Eq. (3.29) $A = 1.1 \times 35.1/9.7 = 4.0$ ft^2
so diameter = 2.26 ft

Overall efficiency

The procedure in Table A-1 is used with the following values:

Diffusivity in the vapor phase, $D_V = 1.73$ ft^2/hr
Gas density, $\rho_G = 0.037$ lb/ft^3
Gas viscosity, $\mu_V = 294.3 \times 10^{-4}$ lb/(ft.hr)

Eq. (11-3.2) in Reference 17 may be used to estimate the diffusivity. This reference and others[28,43] contain data and estimation methods for a number of physical properties.

The above values yield:

$S_c = 0.46$, and
$(NTU)_V = 2.23$

The liquid residence time on a tray of diameter 2.26 ft is:

$$t_L = 0.66 \times 2.26 = 1.5 \text{ seconds}$$

Diffusivities in the liquid phase may be estimated from Eq. (11-10.1) in Reference 17. Using a value of $D_L = 1.67 \times 10^{-4}$ ft^2/hr gives the number of liquid phase transfer units to be:

$$(NTU)_L = 2.61$$

Consequently the number of transfer units is:

$$NTU = (1/2.23 + 1.86/2.61)^{-1}$$
$$= 0.86$$

The point efficiency is:

$$D_E = (10.03 + 1.03 \times 9.7)^2 = 401 \text{ ft}^2/\text{hr}$$

If we take the distance traveled on a plate between weirs to be 0.71 D, or 1.6 ft, then the Peclet number becomes

$$Pe = 3600 \times 1.6^2/(401 \times 1.5) = 15.3$$

In obtaining the Murphree tray efficiency we get:

$M = 1.012$, and
$E_M/E_P = 1.633$, so $E_M = 0.947$

Correcting this efficiency for entrainment using Figure A-1, we get $\Psi = 0.01$ and

$E_A = 0.938$

The overall tower efficiency therefore becomes

$E_o = 0.953$

The actual number of trays is, therefore,

$N = 16/0.95 = 17$ without reflux
$N = 8.7/0.95 = 9$ with reflux

If a reboiler is used, one stage can be subtracted from the actual tray count

2. PACKED TOWER (for 2 inch ceramic packing)

Tower Diameter

From Figure A-1, for abscissa = 0.2, get ordinate = 0.09 at flooding. For a packing factor $C_F = 65$ and a liquid viscosity of 0.29 cP, we get for the vapor flux,

$G' = 1282$ lb/(hr.ft^2) (at flood)
or $G' = 1026$ lb/(hr.ft^2) (at 80% of flood)

Therefore:

Area = 4680/1026 = 4.56 ft^2, and
$D = 2.5$ ft

Number of Transfer Units

From Eq. (3.34), using the same values as above for tray towers,

$$NTU = 21.5 \quad \text{without reflux}$$
$$NTU = 11.7 \quad \text{with reflux}$$

Height of a Transfer Unit and Tower Height

Using the equations in Table 3-5 with viscosity, density and surface tension factors of unity, we get:

$$(HTU)_V = 0.66\ (Z)^{1/3}, \text{ and}$$

$$(HTU)_L = 0.32\ (Z)^{0.15}$$

so, $HTU = Z/NTU = (HTU)_V + S(HTU)_L$

and $Z/21.5 = 0.66\ Z^{1/3} + 0.60\ Z^{0.15}$ for no reflux

Solving for Z gives a packing height of 90 feet.

APPENDIX C

ADIABATIC AIR STRIPPER: BASIC PROGRAM

This program solves the material balance equations in Table 3-2 (no-condenser case) and the enthalpy balance, Eq. (3.20), to determine the number of theoretical stages required for an adiabatic air stripper.

The procedure involves an iteration within an iteration. The temperature of the air leaving the stripper is estimated, and then the enthalpy balance applied to each stage until the bottom stage is reached. If the air temperature entering this stage is not equal to the specified air temperature, a new top temperature must be assumed. The other iteration is made around each stage. The temperature on stage n+1 must be estimated to calculate the enthalpy of the air entering stage n. Eq. (3.20) is then tested, and a new temperature for n+1 estimated until the equation balances.

The program typically runs for several minutes. The use of refined convergence procedures should reduce this time significantly.

ADIABATIC AIR STRIPPER: PROGRAM LISTING

```
 10 GOTO 200
 15 REM   ***   BASIC PROGRAM FOR ADIABATIC AIR STRIPPING   ***
 20 REM   ENTER THE VAPOR-LIQUID CONSTANT K  HERE
 25 REM  WHERE K=y/x  - concentrations are in mole fractions
 30 REM               -  K is dimensionless
 35 REM   (K IS THE HENRY LAW CONSTANT DIVIDED BY PRESSURE)
 40 REM       THE FOLLOWING EQUATIONS MAY BE USEFUL:
 45 REM   LOG (K*P) = LOG (VAP.PRESS) + LOG (GAMMA)
 50 REM   LOG(VAP.PRESS) = A - B/(T + C)
 55 REM   LOG (GAMMA) = D + E/(T + 273)
 60 REM       THE FOLLOWING EQUATION IS FOR TCE
 65 K = 10^(7.03 - 1315/(T + 230) + 1134/(T + 273))/(P * 760)
 70 RETURN
 75 REM   *  EQUATION FOR ENTHALPY OF HUMID AIR, AIRHT, cal/mole
                                                     dry air
 80 AIRHT = 6.96*T + HUMID * (10750 + 8.1*T):RETURN
 85 REM   *  EQUATION FOR MOLE FRACTION WATER VAPOR IN AIR, YW
 90 YW = EXP(21.158 - 5920.8/(T+273) - .006977 * (T+273))/(14.7 * P)
 95 REM   *  EQUATION FOR HUMIDITY OF AIR, HUMID, moles water/mole
                                                       dry air
 100 HUMID = YW/(1 - YW):RETURN
 105 REM   *  EQUATION FOR ENTHALPY OF WATE, WATHT, cal/mole
 110 WATHT = 18 * T:RETURN
 200 DIM L(20),V(20),HUM(20),HEATW(20),HEATA(20),TEMP(20)
 205 DIM Y(20), X(20)
 210 REM  ****   TABLE OF VARIABLES
 215 REM   * CONCENTRATION IN LIQUID, X, IS MOLE FRACTION
 220 REM   * CONCENTRATION IN VAPOR,  Y, IS MOLE RATIO (PER MOLE
                                                     DRY AIR)
 225 REM   * FLOWS ARE IN MOLES/TIME FOR A FEED OF 100 MOLES/TIME
 230 REM L IS LIQUID RATE, MOLES/TIME
 235 REM AIR IS WET AIR RATE, VAIR IS THE DRY AIR RATE
 240 REM N IS STAGE NUMBER, NB IS BOTTOM STAGE
 245 REM   TMAX,TMIN, TTMAX, TTMIN, TWMAX,TWMIN, TNMAX, TNMIN
 250 REM   ARE THE MAX. & MIN. TEMPS FOR THE
 255 REM   TOWER, TOP, WET BULB, AND PLATE N, RESPECTIVELY
 260 REM YW IS MOLE FRACTION WATER VAPOR IN SATURATED AIR
 265 REM YWIN IS MOLE FRACTION WATER VAPOR IN ENTERING AIR
 270 REM YWIN IS MOLE FRACTION WATER VAPOR IN ENTERING AIR
 275 REM TEMP(N) IS TEMPERATURE ON PLATE N
 280 REM EVAP IS THE WATER EVAPORATED ON A STAGE
 285 REM TERROR, HERROR, YERROR, ARE THE
 290 REM   ERRORS ON THE TEMPERATURE, HUMIDITY, AND MOLE FRACTION
                                                  WATER VAPOR
 295 REM HUM(N) IS THE AIR HUMIDITY, MOLES WATER/MOLE DRY AIR
 300 REM HEATW IS THE ENTHALPY OF WATER, CAL/MOLE WATER
 305 REM HEATA IS THE HUMID AIR ENTHALPY, CALS/MOLE DRY AIR
 310 REM ALL TEMPERATURES ARE DEGREES CELSIUS
 315 REM SEE LINES 420 TO 450 FOR INPUT VARIABLES
 400 REM   ****  SPECIFY INPUT VARIABLES  ****
```

```
405 CLS:PRINT"ADIABATIC AIR STRIPPER "
410 PRINT"BEFORE STARTING, YOU SHOULD ENTER THE K-FACTOR"
415 PRINT "  K = y/x IN LINES 20 TO 65"
420 PRINT"ENTER MOLE FRACTION ORGANIC IN LIQUID FEED  ";:INPUT;XF
425 PRINT"ENTER DESIRED CONCENTRATION IN TREATED WATER";:INPUT;XB
430 PRINT"ENTER FEED TEMPERATURE IN CENTIGRADE        ";:INPUT;TF
435 PRINT"ENTER AIR RATE AS MOLE % OF LIQUID FEED     ";:INPUT;AIR
440 PRINT"ENTER TEMPERATURE OF ENTERING AIR           ";:INPUT;TDRY
445 PRINT"ENTER RELATIVE HUMIDITY OF ENTERING AIR, %  ";:INPUT;RH
450 PRINT"ENTER MEAN PRESSURE IN TOWER, ATMOSPHERES   ";:INPUT;P
455 REM    *****   DETERMINE INITIAL VARIABLES  *******
460 L(0) = 100:TEMP(0) = TF:X(0) = XF
465 TMAX = TF:IF TDRY > TF THEN TMAX = TDRY:REM MAX. POSSIBLE
                                                   TEMP IN TOWER
470 TMAX = TMAX + .2
475 TTMAX=TMAX:REM LIMIT FOR TOP TEMPERATURE
480 M = 0:REM COUNT ON TOWER ITERATIONS
485 LN = 0:REM LAST STAGE CALCS
490 REM    ** CALCULATE DRY AIR RATE FROM SAT. HUMIDITY AND
                                                   REL. HUMIDITY
500 T = TDRY:GOSUB 90:HUMIDITY = HUMID * RH/100
505 YWIN = HUMIDITY/(HUMIDITY + 1):VAIR = AIR * (1 - YWIN):
                                       PRINT"DRY AIR RATE = ";VAIR
510 REM   **  CALCULATE THE FEED ENTHALPY (ASSUMING PURE WATER)
515 T = TF:GOSUB 110:FEEDHT = WATHT
520 REM   **  CALCULATE THE WET BULB TEMPERATURE OF THE ENTERING
                                                             AIR
525 IF RH = 100 THEN TWET = TDRY:GOTO 585
530 TWMAX = TDRY:TWMIN = -10
535 TWET = TDRY - 10:REM INITIAL GUESS, ITERATION FOLLOWS:
540 YTEST = YWIN:IF YWIN <9.999999E-06 THEN YTEST = 9.999999E-06
545 T = TWET:GOSUB 90
550 YW2 = YW - (1 - YW)*(TDRY -TWET)/(1555 - .7 * TWET):
                                                REM CARRIER EQN.
555 PRINT"TWET";TWET,"CALC";YW2,YWIN
560 YERROR = (YTEST - YW2)/YTEST
565 IF ABS(YERROR *100) < 1 THEN 585
570 IF YERROR > 0 THEN TWMIN = TWET:GOTO 580
575 IF YERROR < 0 THEN TWMAX = TWET
580 TWET = (TWMIN + TWMAX)/2:GOTO 545
585 TMIN = TWET - .2:IF TF < TWET THEN TMIN = TF - .2:
                                         REM MINIMUM TOWER TEMP.
590 TTMIN=TWET + (TF - TWET)/4:REM LIMIT ON TOP TEMPERATURE
595 IF TWET > TF THEN TTMIN = TF
600 REM    ****   ESTIMATE TOP TEMPERATURE FOR FIRST TIME
605 TEMP(1) = TF - (TF - TWET)/5
610 REM START OF TOWER/TOP TEMP ITERATION
615 REM   **  CALCULATE THE HUMIDITY OF THE LEAVING AIR
620 LN = 0:T = TEMP(1):GOSUB 90:HUM(1) = HUMID :
                               PRINT"TOP HUMIDITY = ";HUM(1)
625 REM   **  CALCULATE THE ENTHALPY OF THE LEAVING AIR
630 GOSUB 80:HEATA(1) = AIRHT
635 REM   **  CALCULATE BOTTOMS WATER RATE
640 REM WATER ADDED TO SATURATE AIR IS
                               VAIR(INITIAL - FINAL HUMIDITY)
```

```
645 WATER = VAIR * (HUM(1) - HUMIDITY)
650 BOTTOMS = 100 - WATER
655 REM   **  CALCULATE THE ORGANIC IN THE AIR LEAVING THE TOWER
660 Y(1) = (100 * XF -BOTTOMS * XB)/VAIR
665 PRINT"BOTTOMS =";BOTTOMS:PRINT"TOP VAPOR = ";Y(1)
670 PRINT"TOP TEMP =";TEMP(1)
700 REM     ***    START OF STAGE CALCULATIONS
705 N=O
710 L(0) = 100
715 N=N+1:IF N = 20 THEN PRINT "Increase air rate":STOP
720 PRINT:PRINT:PRINT N
725 REM ** ENTHALPY BALANCE TO DETERMINE TEMPERATURE ON NEXT PLATE
730 TEMP(N+1) = TEMP(N) - (TEMP(N) - TDRY)/3:REM INITIAL GUESS
735 TNMAX = TMAX:TNMIN = TMIN
740 REM   **  CALCULATE ENTHALPIES AND HUMIDITIES FOR
745 REM   F.FEEDHT + VAIR.AIRHT(N+1) =
                                   VAIR.AIRHT(1) + L(N).WATHT(N)
750 T = TEMP(N):GOSUB 110:HEATW(N) = WATHT
755 T = TEMP(N+1):GOSUB 90:HUM(N+1) = HUMID
760 IF N = NB THEN HUM(N+1) = HUMID * RH/100:PRINT"BOTTOM"
765 HUMID = HUM(N+1):GOSUB 80:HEATA(N+1) = AIRHT
770 REM   **  CALCULATE THE WATER EVAPORATED ON STAGE N
775 EVAP = (HUM(N) - HUM(N+1)) * VAIR:L(N) = L(N-1) - EVAP
780 REM   **  CALCULATE AIR ENTHALPY ON (N+1) BY HEAT BALANCE
785 AIRHT = (VAIR * HEATA(1) + L(N) * HEATW(N) - 100 * FEEDHT)
                                                        /VAIR
790 HERROR = (HEATA(N+1) - AIRHT)/HEATA(N+1) * 100
795 PRINT TEMP(N+1),"ERROR= ";HERROR;"  ON STAGE ";N
800 IF ABS(HERROR) < .01 THEN 860
805 IF HERROR < 0 THEN TNMIN = TEMP(N+1):GOTO 815
810 IF HERROR > 0 THEN TNMAX = TEMP(N+1)
815 TEMP(N+1) = (TNMAX + TNMIN)/2
820 IF (TEMP(N+1) - TMIN) < .01 THEN 835
825 IF (TMAX - TEMP(N+1)) < .01 THEN 840
830 GOTO 750
835 TTMIN = TEMP(1) - (TEMP(1) - TTMIN)/3:GOTO 845
840 TTMAX = TEMP(1) + (TEMP(1) - TTMIN)/3
845 TEMP(1) = (TTMAX + TTMIN)/2
850 PRINT:PRINT "TRYING A NEW TOP TEMPERATURE "
855 GOTO 620
860 REM   ****    CALCULATE THE COMPOSITION ON THE NEXT STAGE
865 REM   Y(N+1) = (VAIR * y(1) + L(N) * x(N) - F * xF)/VAIR
870 REM   **  CALCULATE EQM. LIQUID CONCENTRATION      x = y/K
875 T = TEMP(N):GOSUB 55
880 X(N) = Y(N)/((Y(N) + 1) * K)
885 IF X(N) > XF AND M > 0 THEN PRINT
                         "Increase air rate, x(n)=";X(N):STOP
890 IF NB > 0 THEN 905
895 REM   * CHECK IF LIQUID CONCENTRATION IS LESS THAN BOTTOMS
900 IF X(N) <= XB THEN NB = N: GOTO 730
905 IF N = NB THEN 935
910 REM   **  MATERIAL BALANCE FOR Y(N+1)
915 Y(N+1) = (VAIR * Y(N) + L(N) * X(N) - 100 * XF)/VAIR
920 IF Y(N+1) < 0 THEN Y(N+1) = 0
925 PRINT Y(N+1),Y(N),X(N),K
```

```
930 GOTO 715
935 IF X(N) <= XB THEN 1000
940 NB = NB + 1
945 TTMAX = TTMAX + (TMAX - TTMAX)/5
950 GOTO 620
1000 REM   **  CHECK INITIAL TEMPERATURE GUESS
1005 TERROR = (TDRY - TEMP(N+1))/TDRY
1010 IF ABS(TERROR) < 5.000001E-03 THEN 1200
1015 M = M + 1
1020 PRINT N,TEMP(N+1),TDRY,TERROR
1025 PRINT:PRINT:PRINT
1030 IF TERROR < 0 THEN TTMAX = TEMP(1):GOTO 1040
1035 IF TERROR > 0 THEN TTMIN = TEMP(1)
1040 TEMP(1) = (TTMAX + TTMIN)/2:GOTO 1045
1045 IF M > 20 THEN 1200
1050 GOTO 620
1200 REM ** PRINT RESULTS TO SCREEN  ** Modify format as required.
1205 CLS
1210 PRINT"INPUT DATA"
1215 PRINT"FEED CONC=";XF;"  FEED T=";TF;"  BOTTOMS=";XB;"  K=";K
1220 PRINT"AIR RATE=";AIR;"  AIR TEMP= ";TDRY;"  PRESSURE=";P;
                                                          "RH=";RH
1225 PRINT:PRINT"CALCULATED VALUES"
1230 PRINT"TWET = ";TWET;"   BOTTOMS = ";BOTTOMS
1235 PRINT"CALC AIR HUM= ";HUM(N+1);"  GIVEN = ";HUMIDITY
1240 PRINT:PRINT"STAGE      TEMP         LIQUID    X*1000      Y*1000"
1245 PRINT"FEED       ";:PRINT USING "####.###   ";TEMP(0),L(0),X(0)
                                                               *1000
1250 FOR I = 1 TO N:PRINT I;"       ";:PRINT USING "####.###   ";
                           TEMP(I),L(I),X(I)*1000,Y(I)*1000:NEXT I
1255 PRINT "AIR        ";TEMP(N+1);" (GIVEN AS ";TDRY;
                              "   ERROR % =";-TERROR * 100;")"
1260 IF X(NB-1) < XB THEN PRINT:PRINT
              " Number of stages falls between ";NB;" and ";NB-1
1270 STOP
```

TYPICAL OUTPUT

```
INPUT DATA
FEED CONC= .1   FEED T= 40   BOTTOMS= .0000001   K= 578.588
AIR RATE= 75   AIR TEMP=  20   PRESSURE= 1   RH= 50

CALCULATED VALUES
TWET =  13.90625    BOTTOMS =  97.2744
CALC AIR HUM=  1.159221E-02   GIVEN =  1.165858E-02

STAGE      TEMP        LIQUID    X*1000      Y*1000
FEED        40.000     100.000   100.000
 1         31.916     98.912      0.169    134.888
 2          25.935     98.179      0.000      0.225
 3          20.368     97.269      0.000      0.000
AIR         19.91004  (GIVEN AS  20    ERROR % =-.449791 )
```

APPENDIX D

ANTOINE COEFFICIENTS: BASIC PROGRAM

The program calculates the Antoine coefficients **A**, **B**, **C** in Eq. (2.15) from three pairs of temperature-vapor pressure data. The data may be in mixed units. The coefficients calculated are for determining vapor pressure in atmospheres using temperature in Kelvin.

PROGRAM LISTING

```
10 REM        Program to calculate the Antoine coefficients
20 REM
30 REM          Antoine equation: log(10)p = A - B/(C + T)
40 REM           p is vapor pressure, T is temperature.
50 REM           Units are program selectable.
60 REM
70 REM  *  Requires 3 vapor pressure - temperature data points *
80 REM   Conversion factors for pressure and temperature
90 DIM PVERT(5),TVERT(3,3)
100 REM   Pressure and temperature data points
110 REM   `Subscript' IN for input value, CALC for calculated value,
120 REM    CON for input value converted to selected units
130 REM     and ERR is for error between input and calculated value.
140 DIM PIN(3),PCALC(3),PCON(3),PERR(3),TIN(3),TCON(3)
150 DIM ALP(3):REM log of vapor pressure
160 REM Alpha variables for units
170 DIM PUNIT$(5),TUNIT$(3)
180 REM        Read units
190 DATA "mm Hg   ", "inch Hg ", "psi     ", "bar     ", "atm     "
200 DATA "F       ", "C       ", "K       "
210 FOR I=1 TO 5:READ Z$:PUNIT$(I)=Z$:NEXT I
220 FOR I=1 TO 3:READ Z$:TUNIT$(I)=Z$:NEXT I
230 REM     Read conversion factors
240 REM Pressure: from mm Hg, inch Hg, psi, bar, to atm
250 DATA 760, 29.92,14.696,0.9872,1.00
260 REM Temperature:from F, C, to Kelvin
270 DATA 0.5556,255.37,1.0,273.15,1.0,0
280 FOR I=1 TO 5
290 READ Z:PVERT(I)=1/Z
300 NEXT I
310 FOR I=1 TO 3:FOR J=1 TO 2
320 READ Z:TVERT(I,J)=Z
330 NEXT J:NEXT I
500 CLS
510 PRINT"Calculate coefficients in the Antoine Equation"
511 PRINT"     log10 p(atm) = A - B/[C + T(Kelvin)]"
512 PRINT"Any of the following units may be used in mixed mode"
```

```
513 PRINT"The data will be converted to atm. and Kelvin"
530 PRINT"Pressure units:";
540 FOR I=1 TO 5:PRINT I; "=";PUNIT$(I);:NEXT I
550 PRINT"Temperature   :";
560 FOR I=1 TO 3:PRINT I; "=";TUNIT$(I);:NEXT I
565 PRINT:PRINT"Data entry":
570 FOR N=1 TO 3
580 PRINT "PRESSURE,    unit #";:INPUT; I:IF I> 5 THEN GOTO 580
590 PRINT" Value";:INPUT;PIN(N):PRINT PUNIT$(I),
600 PCON(N)=PIN(N) * PVERT(I)
610 IF I < 5 THEN PRINT " or ";PCON(N); PUNIT$(5);
620 PRINT:PRINT "TEMPERATURE, unit #";:INPUT; I:IF I>3 THEN GOTO 620
630 PRINT" Value";:INPUT;TIN(N):PRINT TUNIT$(I),
640 TCON(N)=TIN(N) * TVERT(I,1)+TVERT(I,2)
650 IF I < 3 THEN PRINT " or ";TCON(N); TUNIT$(3);
660 PRINT:PRINT:NEXT N
710 FOR N=1 TO 3
720 ALP(N)=LOG(PCON(N))/2.302585
730 NEXT N
740 D1=(TCON(1)-TCON(2))/(ALP(2)-ALP(1))
750 D2=(TCON(2) - TCON(3))/(ALP(3) - ALP(2))
760 D3=(ALP(3) * TCON(3) - ALP(2) * TCON(2))/(ALP(3) - ALP(2))
770 D4=(ALP(2) * TCON(2) - ALP(1) * TCON(1))/(ALP(2) - ALP(1))
780 A=(D3 - D4)/(D1 - D2)
790 C=-1 * (A * D1) - D4
800 B=(A-ALP(1)) * (C + TCON(1))
805 PRINT:PRINT"Input T, K    Input P, atm    Calc. P,    % error:"
810 FOR N=1 TO 3
820 PCALC(N) = 10^(A - B/(C+TCON(N)))
830 PERR(N)=(PCON(N) - PCALC(N)) * 100/PCON(N)
840 IF PERR(N) < .001 THEN PERR(N) = 0
860 PRINT"  ";
870 PRINT USING "###.###       ";TCON(N);PCON(N);PCALC(N);PERR(N)
880 NEXT N
890 PRINT:PRINT"A =";A;"  B =";B;"  C =";C
```

APPENDIX E

TOXIC ORGANICS LIST

Data from Reference 3 is summarized here for convenience. For a method of predicting activity coefficients and Henry's Law constant for compounds not listed here, see Reference 45. Vapor pressure data are listed in Reference 46.

Compounds directory

Ethers	5-1 to 5-7
Phthalates	6-1 to 6-6
Nitrogen Compounds	7-1 to 7-14
Phenols	8-1 to 8-14
Aromatics	9-1 to 9-20
Polynuclear Aromatic Hydrocarbons	10-1 to 10-17
PCB's and Related Compounds	11-1 to 11-7
Halogenated Hydrocarbons	12-1 to 12-31
Pesticides	13-1 to 13-46
Oxygenated Compounds	14-1 to 14-18
Miscellaneous	15-1 to 15-7

Stripping Category, S

1	Very easily stripped
2	Easily stripped
3	Intermediate
4	Difficult
5	Very difficult
6	Cannot be stripped
D	Decomposes in water
P	Poor data, uncertain classification

EASE OF STRIPPING, HENRY'S LAW CONSTANT AND ACTIVITY COEFFICIENTS

5-1 Bis(chloromethyl)ether S = D
log H(mm Hg) = decomposes
log (Activity coefficient) = decomposes

5-2 Bis(2-chloroethyl)ether S = 3
$\log H(\text{mm Hg}) = 7.659 - 1969/(t^{\circ}C + 234) + 862/(t^{\circ}C + 273)$
$\log(\text{Activity coefficient}) = 862/(t^{\circ}C + 273)$

5-3 Bis(2-chloroisopropyl)ether S = 2
$\log H(\text{mm Hg}) = 7.94 - 1231/(t^{\circ}C + 273)$
$\log(\text{Activity coefficient}) = 1117/(t^{\circ}C + 273)$

5-4 2-Chloroethyl vinyl ether S = 1
$\log H(\text{mm Hg}) = 7.75 - 1079/(t^{\circ}C + 273)$
$\log(\text{Activity coefficient}) = 774/(t^{\circ}C + 273)$

5-5 4-Chlorophenyl phenyl ether S = 4-P
$H(100^{\circ}C) = 3.9$ atm
$\text{Activity coefficient}(100^{\circ}C) = 2.948 \times 10^{3}$

5-6 4-Bromophenyl phenyl ether S = 3-P
$H(100^{\circ}C) = 9.9$ atm
$\text{Activity coefficient}(100^{\circ}C) = 4418$

5-7 Bis(2-chloroethoxy)methane S = 6
$\log H(\text{mm Hg}) = 8.151 - 2416/(t^{\circ}C + 243) + 618/(t^{\circ}C + 273)$
$\log(\text{Activity coefficient}) = 618/(t^{\circ}C + 273)$

6-1 Dimethyl phthalate S = 6
$\log H(\text{mm Hg}) = 8.293 - 2858/(t^{\circ}C + 244) + 1036/(t^{\circ}C + 273)$
$\log(\text{Activity coefficient}) = 1036/(t^{\circ}C + 273)$

6-2 Diethyl phthalate S = 5

log H(mm Hg) = 8.29 - 2883/($t^{o}C$ + 239) + 1248/($t^{o}C$ + 273)

log (Activity coefficient) = 1248/($t^{o}C$ + 273)

6-3 Di-n-butyl phthalate S = 4-P

H(25°C) = 0.15 atm

Activity coefficient(25°C) = 1.2×10^{6}

6-4 Di-n-octylphthalate S = 3-P

H(100°C)) = 9,947 atm

Activity coefficient(100°C) = 5.4×10^{7}

6-5 Bis(2-ethylhexyl) phthalate S = 4-P

log H(mm Hg) = data uncertain

log (Activity coefficient) = data uncertain

6-6 Butyl benzyl phthalate S = 3-P

log H(mm Hg) = data uncertain

log (Activity coefficient) = data uncertain

7-1 N-nitrosodimethylamine S = 4-P

log H(mm Hg) = no data

log (Activity coefficient) = no data

7-2 N-nitrosodiphenylamine S = 4-P

log H(mm Hg) = no data

log (Activity coefficient) = no data

7-3 N-nitrosodi-n-propylamine S = 3-P

log H(mm Hg) = data uncertain

log (Activity coefficient) = data uncertain

7-4 Benzidine S = 3-P

log H(mm Hg) = data uncertain

log (Activity coefficient) = data uncertain

7-5 3,3'-Dichlorobenzidine S - 4-P
H(100°C)) = 0.013 atm
Activity coefficient(100°C) = 65

7-6 1,2-Diphenylhydrazine S = 4-P
H(100°C)) = 0.006 atm
Activity coefficient(100°C) = 19

7-7 Acrylonitrile S = 2
log H(mm Hg) = 6.655 - 1208/(t°C + 222) + 545.6/(t°C + 273)
log (Activity coefficient) = (545.6/T) - 0.2606

7-8 n-Butylamine S = P
log H(mm Hg) = no data
log (Activity coefficient) = no data.

7-9 Diethylamine S = 3-P
log H(mm Hg) = 16.52 - 1127/(t°C + 220) - 2804/(t°C + 273)
log (Activity coefficient) = 9.55 - 2804/(t°C + 273)

7-10 Ethylene diamine S = 6
log H(mm Hg) = no data
log (Activity coefficient) = no data

7-11 Monoethylamine S = P
log H(mm Hg) = no data
log (Activity coefficient) = no data

7-12 Monomethylamine S = P
log H(mm Hg) = no data
log (Activity coefficient) = no data

7-13 Triethylamine S = 1
H(20°C) = 26 atm
Activity coefficient(20°C) = 375

7-14 Trimethylamine S = P
log H(mm Hg) = no data
log (Activity coefficient) = no data

8-1 Phenol S = 5
log H(mm Hg) = 9.071 - 0.00352 t°C - 1517/(t°C + 174)
log(Activitycoefficient)= 1.941-0.00352 t°C(55 to 240°C)

8-2 2-Chlorophenol S = 4
log H(mm Hg) = 7.24 - 1668/(t°C + 210) + 715/(t°C + 273)
log(Activitycoefficient)= 715/(t°C + 273)

8-3 2,4-Dichlorophenol S = 4
log H(mm Hg) = 8.205 - 2380/(t°C + 237) + 992/(t°C + 273)
log(Activitycoefficient)= 992/(t°C + 273)

8-4 2,4,6-Trichlorophenol S = 4
log H(mm Hg) = 8.096 - 2484/(t°C + 230) + 1230/(t°C + 273)
log(Activitycoefficient)= 1230/(t°C + 273)

8-5 Pentachlorophenol S = 3
log H(mm Hg) = 12.047 - 7751/(t°C + 509) + 1800/(t°C + 273)
log(Activitycoefficient)= 1800/(t°C + 273)

8-6 2-Nitrophenol, O-nitrophenol S = 3
log H(mm Hg) = 8.014 - 2173/(t°C + 231) + 971/(t°C + 273)
log(Activitycoefficient)= 0.251 + 971/(t°C + 273)

8-7 4-Nitrophenol, p-nitrophenol S = 6
log H(mm Hg) = 7.483 - 2604/(t°C + 273)
log(Activitycoefficient)= -3.412 + 1816/(t°C + 273)

8-8 2,4-Dinitrophenol S = 6
log H(mm Hg) = no data
log (Activity coefficient) = -0.767 + 1173/(t°C + 273)

8-9 Resorcinol

S = 6

log H(mm Hg) = 5.31 - 3049/(t°C + 238) + 1191/(t°C + 273)

log(Activitycoefficient)= -3.50 + 1191/(t°C + 273)

8-10 2,4-Dimethyl phenol, 2,4 Xylenol

S = 5

log H(mm Hg) = 7.06 - 1587/(t°C + 170) + 923/(t°C + 273)

log(Activitycoefficient)= 923/(t°C + 273)

8-11 Total phenols

See individual phenols

8-12 p-chloro-m-cresol

S = 3-P

H(100°C) = 8.2 atm

Activity coefficient(100°C) = 376

8-13 4,6-Dinitro-o-cresol

S = 5

log H(mm Hg) = uncertain data

log (Activity coefficient) = no data

8-14 Cresols

S = 5

log H(mm Hg) = 7.51 - 1856/(t°C + 199) + 760/(t°C + 273)

log(Activitycoefficient)= 760/(t°C + 273)

9-1 Benzene

S = 1

log H(mm Hg) = 7.455 - 1211/(t°C + 221) + 848.2/(t°C + 273)

log(Activitycoefficient)= 0.5455 + 848.2/(t°C + 273)

9-2 Chlorobenzene

S = 1

log H(mm Hg) = 6.845 - 1431/(t°C + 218) + 1274/(t°C + 273)

log(Activitycoefficient)= -0.135 + 1274/(t°C + 273)

9-3 1,2-Dichlorobenzene, o-Dichlorobenzene

S = 1

log H(mm Hg) = 7.07 - 1650/(t°C + 213) + 1421/(t°C + 273)

log(Activitycoefficient)= 1421/(t°C + 273)

9-4 1,3-Dichlorobenzene, m-Dichlorobenzene S = 1
log H(mm Hg) = 7.30 - 1732/(t°C + 230) + 1407/(t°C + 273)
log(Activitycoefficient)= 1407/(t°C + 273)

9-5 1,4-Dichlorobenzene, p-Dichlorobenzene S = 1
log H(mm Hg) = 7.00 - 1575/(t°C + 209) + 1425/(t°C + 273)
log(Activitycoefficient)= 1425/(t°C + 273)

9-6 1,2,4-Trichlorobenzene S = 1
log H(mm Hg) = 7.601 - 2175/(t°C + 248) + 1533/(t°C + 273)
log(Activitycoefficient)= 1533/(t°C + 273)

9-7 Hexachlorobenzene S = 1
log H(mm Hg) = 9.836 - 4630/(t°C + 356) + 2138/(t°C + 273)
log(Activitycoefficient)= 2138/(t°C + 273)

9-8 Ethyl Benzene S = 1
log H(mm Hg) = 6.286 - 1424/(t°C + 213) + 1529/(t°C + 273)
log(Activitycoefficient)= -0.664 + 1529/(t°C + 273)

9-9 Nitrobenzene S = 3
log H(mm Hg) = 6.856 - 1740/(t°C + 200) + 1141/(t°C + 273)
log(Activitycoefficient)= -0.262 + 1141/(t°C + 273)

9-10 Toluene S = 1
log H(mm Hg) = 6.95 - 1345/(t°C + 219) + 1192/(t°C + 273)
log(Activitycoefficient)= -0.0004 + 1192/(t°C + 273)

9-11 2,4-Dinitrotoluene S = 3
log H(mm Hg) = 7.450 - 1238/(t°C + 273)
log(Activitycoefficient)= 1380/(t°C + 273)

9-12 2,6-Dinitrotoluene S = 3
Use data for 9-11

9-13 Aniline S = 6

See Reference 3

9-14 Benzoic acid S = 6

$\log H(\text{mm Hg}) = 9.922 - 2816/(t^{\circ}\text{C} + 273)$

$\log(\text{Activitycoefficient}) = 0.8894 + 516.6/(t^{\circ}\text{C} + 273)$

9-15 Benzyl chloride S = 1

$\log H(\text{mm Hg}) = 7.546 - 1923/(t^{\circ}\text{C} + 233) + 1265/(t^{\circ}\text{C} + 273)$

$\log(\text{Activitycoefficient}) = 1265/(t^{\circ}\text{C} + 273)$

9-16 Styrene S = 1

$\log H(\text{mm Hg}) = 9.31 - 1446/(t^{\circ}\text{C} + 209) + 570/(t^{\circ}\text{C} + 273)$

$\log(\text{Activitycoefficient}) = 2.34 + 570/(t^{\circ}\text{C} + 273)$

9-17 Quinoline S = 5

$\log H(\text{mm Hg}) = 7.89 - 2398/(t^{\circ}\text{C} + 244) + 619/(t^{\circ}\text{C} + 273)$

$\log(\text{Activitycoefficient}) = 619/(t^{\circ}\text{C} + 273)$

9-18 Xylenes S = 1

$\log H(\text{mm Hg}) = 7.00 - 1460/(t^{\circ}\text{C} + 215) + 1339/(t^{\circ}\text{C} + 273)$

$\log(\text{Activitycoefficient}) = 1339/(t^{\circ}\text{C} + 273)$

9-19 Nitrotoluene S = 2

Ortho: $\log H(\text{mm Hg}) = 7.97 - 1263/(t^{\circ}\text{C} + 273)$

meta: $\log H(\text{mm Hg}) = 8.07 - 1368/(t^{\circ}\text{C} + 273)$

para: $\log H(\text{mm Hg}) = 7.98 - 1359/(t^{\circ}\text{C} + 273)$

$\log(\text{Activity coefficient}) = 1250/(t^{\circ}\text{C} + 273)$

9-20 Napthenic acid

No data

10-1 2-Chloronaphthalene S = 1

$\log H(\text{mm Hg}) = 9.93 - 1644/(t^{\circ}\text{C} + 273)$

$\log(\text{Activity coefficient}) = 1785/(t^{\circ}\text{C} + 273)$

10-2 Benzo(a)anthracene S = 3-P
Uncertain data

10-3 Benzo(b)fluoranthene S = 1-P
$H(100^oC) = 208$ atm
log (Activity coefficient) = $2600/(t^oC + 273)$

10-4 Benzo(k)fluoranthene S = 1-P
log H(mm Hg) = $7.445 - 837/(t^oC + 273)$
log (Activity coefficient) = $2600/(t^oC + 273)$

10-5 Benzo(a)pyrene S = 1-P
log H(mm Hg) = $5.57 + 47/(t^oC + 273)$
log (Activity coefficient) = $2716/(t^oC + 273)$

10-6 Indeno(1,2,3-cd)pyrene S = 3-P
Uncertain data

10-7 Dibenzo(a,h)anthracene S = 3-P
Uncertain data

10-8 Benzo(ghi)porylene S = 6
$H(100^oC) = 0.104$ atm
log (Activity coefficient) = $2946/(t^oC + 273)$

10-9 Acenaphthene S = 1
log H(mm Hg) = $8.033 - 1200/(t^oC + 273)$
log (Activity coefficient) = $1635/(t^oC + 273)$

10-10 Acenaphthylene S = 2
$H(25^oC) = 6.3$ atm
Activity coefficient$(25^o) = 20$

10-11 Anthracene S = 3
log H(mm Hg) = $8.91 - 1739/(t^oC + 273)$
log (Activity coefficient) = $2022/(t^oC + 273)$

10-12 Chrysene S = 1-P

Data uncertain

10-13 Fluoranthene S = 1-P

Data uncertain

10-14 Fluorene S = 2

$\log H(\text{mm Hg}) = 8.06 - 1301/(t^{o}C + 273)$

$\log (\text{Activity coefficient}) = 1632/(t^{o}C + 273)$

10-15 Naphthalene S = 1

$\log H(\text{mm Hg}) = 7.01 - 1734/(t^{o}C + 202) + 1473/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 1473/(t^{o}C + 273)$

10-16 Phenanthrene S = 2

$\log H(\text{mm Hg}) = 7.26 - 2379/(t^{o}C + 204) + 2022/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 2022/(t^{o}C + 273)$

10-17 Pyrene S = 1

H = Data uncertain

$\log (\text{Activity coefficient}) = 2365/(t^{o}C + 273)$

10-18 Aroclor 1016 S = 1

$\log H(\text{atm}) = 6.22 - 1555/(t^{o}C + 273)$

Activity coefficient $(100^{o}C) = 1.0 \times 10^{5}$

11-2 Aroclor 1221 S = 1

$\log H(\text{atm}) = 8.05 - 2026/(t^{o}C + 273)$

Activity coefficient $(100^{o}C) = 1.1 \times 10^{5}$

11-3 Aroclor 1232 S = 1

$\log H(\text{atm}) = 8.24 - 1954/(t^{o}C + 273)$

Activity coefficient $(100^{o}C) = 3.6 \times 10^{5}$

11-4 Aroclor 1242 S = 1
$\log H(\mathrm{atm}) = 10.01 - 2577/(t^{\circ}C + 273)$
Activity coefficient ($100^{\circ}C$) = 1.3×10^{6}

11-5 Aroclor 1248 S = 1
$\log H(\mathrm{atm}) = 9.80 - 2313/(t^{\circ}C + 273)$
Activity coefficient ($100^{\circ}C$) = 5.9×10^{6}

11-6 Aroclor 1254 S = 1
$\log H(\mathrm{atm}) = 9.46 - 2024/(t^{\circ}C + 273)$
Activity coefficient ($100^{\circ}C$) = 2.1×10^{7}

11-7 Aroclor 1260 S = 1
$\log H(\mathrm{atm}) = 9.28 - 1928/(t^{\circ}C + 273)$
Activity coefficient ($100P^{\circ}C$) = 7×10^{7}

12-1 Methyl chloride, chloromethane S = 1
$\log H(\mathrm{mm\ Hg}) = 5.233 + 276/(t^{\circ}C + 273)$
$\log (\text{Activity coefficient}) = -2.247 + 1424/(t^{\circ}C + 273)$

12-2 Methylene chloride, dichloromethane S = 1
$\log H(\mathrm{mm\ Hg}) = 9.58 - 1139/(t^{\circ}C + 231)$
Insufficient knowledge on the variation of the activity coefficient with temperature; a value of 315 (independent of temperature) can be used.

12-3 Chloroform S = 1
$\log H(\mathrm{mm\ Hg}) = 10.67 - 1171/(t^{\circ}C + 227) - 260/(t^{\circ}C + 273)$
$\log (\text{Activitycoefficient}) = 3.73 - 260/(t^{\circ}C + 273)$

12-4 Carbon tetrachloride S = 1
$\log H(\mathrm{mm\ Hg}) = 6.796 - 1220/(t^{\circ}C + 227) + 1237/(t^{\circ}C + 273)$
$\log (\text{Activitycoefficient}) = -0.0979 + 1237/(t^{\circ}C + 273)$

12-5 Chloroethane, ethyl chloride S = 1

$\log H(\text{mm Hg}) = 6.324 + 993/(t^{o}C + 273) - 1013/(t^{o}C + 237)$

$\log (\text{Activitycoefficient}) = -0.616 + 993/(t^{o}C + 273)$

12-6 1,1-Dichloroethane S = 1

$\log H(\text{mm Hg}) = 6.99 - 1171/(t^{o}C + 228) + 895/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 895/(t^{o}C + 273)$

12-7 1,2-Dichloroethane S = 1

$\log H(\text{mm Hg}) = 7.03 - 1271/(t^{o}C + 223) + 824/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 824/(t^{o}C + 273)$

12-8 1,1,1-Trichloroethane S = 1

$\log H(\text{mm Hg}) = 7.03 - 1276/(t^{o}C + 234) + 947/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 947/(t^{o}C + 273)$

12-9 1,1,2-Trichloroethane S = 1

$\log H(\text{mm Hg}) = 7.17 - 1351/(t^{o}C + 217) + 884/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 0.199 + 884/(t^{o}C + 273)$

12-10 1,1,2,2,-Tetrachloroethane S = 1

$\log H(\text{mm Hg}) = 7.54 - 1683/(t^{o}C + 234) + 960/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 0.227 + 960/(t^{o}C + 273)$

12-11 Hexachloroethane S = 1

$\log H(\text{mm Hg}) = 8.649 - 903/(t^{o}C + 273)$

$\log (\text{Activity coefficient}) = 1745/(t^{o}C + 273)$

12-12 Vinyl chloride S = 1

$H\ (100^{o}C) = 4.6\times10^{3}$ atm

Activity coefficient $(100^{o}C) = 145$

12-13 1,2-Dichloropropane S = 1

$\log H(\text{mm Hg}) = 6.966 - 1296/(t^{o}C + 221) + 1010/(t^{o}C + 273)$

$\log (\text{Activitycoefficient}) = 1010/(t^{o}C + 273)$

12-14 1,3-Dichloropropene S = 1

$\log H(\text{atm}) = 4.543 - 670/(t^{\circ}C + 273)$

12-15 Hexachlorobutadiene S = 1

$\log H(\text{mm Hg}) = 5.819 - 1029/(t^{\circ}C + 130) + 1926/(t^{\circ}C + 273)$

$\log (\text{Activitycoefficient}) = 1928/(t^{\circ}C + 273)$

12-16 Hexachlorocyclopentadiene S = 1

$\log H(\text{mm Hg}) = 8.00 - 644(t^{\circ}C + 273)$

$\log (\text{Activity coefficient}) = 2067/(t^{\circ}C + 273)$

12-17 Methyl Bromide S = 1

$\log H(\text{mm Hg}) = 6.96 - 987/(t^{\circ}C + 238) + 1123/(t^{\circ}C + 273)$

$\log (\text{Activitycoefficient}) = -0.00200 + 1123/(t^{\circ}C + 273)$

12-18 Dichlorobromomethane S = 1

$\log H(\text{atm}) = 3.795 - 513/(t^{\circ}C + 273)$

12-19 Chlorodibromomethane S = 1

$\log H(\text{atm}) = 5.262 - 1078/(t^{\circ}C + 273)$

12-20 Bromoform, Tribromomethane S = 1

$\log H(\text{atm}) = 5.464 - 1188/(t^{\circ}C + 273)$

12-21 Dichlorodifluoromethane S = 1

$\log H(\text{mm Hg}) = 5.954 - 382.6/(t^{\circ}C + 145) + 1305/(t^{\circ}C + 273)$

$\log (\text{Activitycoefficient}) = 1305/(t^{\circ}C + 273)$

12-22 Trichlorofluoromethane S = 1

$\log H(\text{mm Hg}) = 6.884 - 1043/(t^{\circ}C + 237) + 1040/(t^{\circ}C + 273)$

$\log (\text{Activitycoefficient}) = 1040/(t^{\circ}C + 273)$

12-23 Trichloroethylene S = 1

$\log H(\text{mm Hg}) = 7.03 - 1315/(t^{\circ}C + 230) + 1134/(t^{\circ}C + 273)$

$\log (\text{Activitycoefficient}) = 1134/(t^{\circ}C + 273)$

12-24 1,1-Dichloroethylene S = 1
$\log H(\text{mm Hg}) = 7.606 - 1471/(t^{\circ}C + 280) + 903/(t^{\circ}C + 273)$
$\log (\text{Activitycoefficient}) = 903/(t^{\circ}C + 273)$

12-25 1,2-Trans-Dichloroethylene S = 1
$\log H(\text{mm Hg}) = 7.037 - 1145/(t^{\circ}C + 228) + 846/(t^{\circ}C + 273)$
$\log (\text{Activitycoefficient}) = 846/(t^{\circ}C + 273)$

12-26 Tetrachloroethylene S = 1
$\log H(\text{mm Hg}) = 6.96 - 1415/(t^{\circ}C + 221) + 1449/(t^{\circ}C + 273)$
$\log (\text{Activitycoefficient}) = -0.061 + 1449/(t^{\circ}C + 273)$

12-27 Allyl Chloride S = 1
$H (\geq 25^{\circ}C) = 20{,}000$ atm
Activity coefficient $(25^{\circ}C) = 42{,}500$

12-28 2,2-Dichloropropionic acid S = P
No data

12-29 Phosgene S = D
Hydrolyzes in water

12-30 Ethylene dibromide, 1,2-Dibromoethane S = 1
$\log H(\text{mm Hg}) = 9.044 - 2989/(t^{\circ}C + 353) + 1024/(t^{\circ}C + 273)$
$\log (\text{Activitycoefficient}) = 1{,}024/(t^{\circ}C + 273)$

12-31 Epichlorohydrin S = 3
$\log H(\text{mm Hg}) = 7.487 - 1609/(t^{\circ}C + 231) + 586/(t^{\circ}C + 273)$
$\log (\text{Activitycoefficient}) = 586/(t^{\circ}C + 273)$

13-1 α-Endosulfan S = P
Data uncertain

13-2 Endosulfan sulfate S = 6
No data

13-3 β-Endosulfan S = P
No data

13-4 α-BHC S = P
Data uncertain

13-5 β-BHC S = P
Data uncertain

13-6 δ-BHC S = P
Data uncertain

13-7 γ-BHC S = P
Data uncertain

13-8 Aldrin S = 2
H (20° to 25°C) = 15 atm
Activity coefficient (20°C) = 1.84×10^9

13-9 Dieldrin S = 2
Data uncertain

13-10 4,4'-DDE S = 3-P
H (20° to 25°C) = 1.3 atm
Activity coefficient (25°C) = 1.5×10^8

13-11 4,4'-DDT S = 6
H (approx. 25°C) = 0.8 to 0.03 atm
Activity coefficient (approx. 25°C) = 1.1×10^8 to 3.3×10^9

13-12 4,4'-DDD S = 3-P
H (25° to 30°C) = 1.2 atm
Activity coefficient (25°C) = 8.9×10^8

13-13 Endrin S = 6
H (25°C) = 0.039 to 0.021 atm
Activity coefficient (25°C) = $1.1x10^{8}$ to $8.1x10^{7}$

13-14 Kelthane S = P
No data

13-15 Naled S = P
No data

13-16 Dichlone S = P
No data

13-17 Kepone S = 4-P
No data

13-18 Diuron S = 6
H (temp. not known) = 0.0013 atm

13-19 Endrin Aldehyde S = P
No data

13-20 Heptachlor S = 1
H (25°C) = 82 atm

13-21 Heptachlor Epoxide S = 3
H (25°C) = 1.8 atm

13-22 Carbofuran S = 6
H (25°C) = $4.6x10^{-4}$ atm
Activity coefficient (25°C) = $1.76x10^{4}$

13-23 Mercaptodimethur S = P
No data

13-24 Chlordane S = 2
Data uncertain

13-25 Toxaphene S = 1
Data uncertain

13-26 Captan S = 2-P
Data uncertain

13-27 Carbaryl S = 3-P
No data

13-28 Coumaphos S = 6
H (20° to 25°C) = 0.0018 atm

13-29 Diazinon S = 6
H (20°C) = 0.078 atm

13-30 Dicamba S = 6
H (100°C) = 0.027 atm

13-31 Dichlobenil S = 6
H (20°C) = 0.28 atm

13-32 Malathion S = 6
H = 0.0067 atm

13-33 Methyl parathion S = 6
H (20° to 25°C) = 0.0034 to 0.0031 atm

13-34 Parathion S = 6
H (20° to 25°C) = 0.034 atm

13-35 Guthion S = 6
H (20° to 25°C) < 0.21 atm

13-36 Ethion S = P
No data

13-37 Isoprene S = 1-P
No data

13-38 Chlorpyridos S = 6
H (25° to 35°C) = 0.23 atm

13-39 Dichlorvos S = 6
H (20°C) = 0.019 atm

13-40 Diquat S = 6-P
No data

13-41 Disalfoton S = 6
H (20° to 23°C) = 0.14 atm

13-42 Mevinphos S = 6-P
No data

13-43 Mexacarbate S = 6
H (139°C) < 0.63 atm
Activity coefficient (139°C) = 4.8x10^{3}

13-44 Trichlorfon S = 6
H (25°C) = 9.5x10^{-7} atm

13-45 Propargite S = P
No data

13-46 Carbon disulfide S = 1
log H(mm Hg) = 6.942 - 1169/(t°C + 242) + 949/(t°C + 273)
log (Activitycoefficient)= 949/(t°C + 273)

14-1 Acetaldehyde S = 2

log H(mm Hg) = 7.293 + 0.00751t°C - 1071/(t°C + 236)

log (Activity coefficient) = 0.4829 + 00751t°C

14-2 Acetic Acid S = 6

log H(mm Hg) = 7.30 - 1479/(t°C + 217) + 133/(t°C + 273)

log (Activitycoefficient)= 133/(t°C + 273)

14-3 Allyl Alcohol S = 5

log H(mm Hg) = 9.97 - 1272/(t°C + 188) - 600/(t°C + 273)

log (Activitycoefficient)= 2.63 - 600/(t°C + 273)

14-4 iso-Amyl Acetate S = 1

log H(atm) = 8.702 - 2187/(t°C + 273)

Activity cosfficient (100°C) = 2730

14-5 n-Butyl Acetate S = 1

log H(mm Hg) = 7.03 - 1368/(t°C + 204) + 965/(t°C + 273)

log (Activitycoefficient)= 965/(t°C + 273)

14-6 Butyric Acid S = 6

H (100°C) = 1 to 2.2 atm

14-7 Formaldehyde S = 6

Data uncertain

14-8 Formic Acid S = 6

No data

14-9 Fumaric Acid S = 4-P

No data

14-10 Maleic Acid S = 4-P

Data uncertain

14-11 Methyl methacrylate S = P
No data

14-12 Propionic Acid S = 6
See Reference 3

14-13 Vinyl Acetate S = 1
log H(mm Hg) = 6.992 - 1192/(t°C + 217) + 697/(t°C + 273)
log (Activitycoefficient)= 697/(t°C + 273)

14-14 Adipic Acid S = 6
log H(mm Hg) = 8.639 - 3074/(t°C + 196) + 787/(t°C + 273)
log (Activitycoefficient)= 787/(t°C + 273)

14-15 Crotononaldehyde S = 4
log H(atm) = 5.24 - 1571/(t°C + 273)
Activity coefficient (20°C) = 22 to 25

14-16 Acrolein S = 2-P
log H(mm Hg) = 8.17 - 1132/(t°C + 228)
log (Activity coefficient) = 1.2634

14-17 Furfural S = 3-P
H (100°C) = 1.3 to 25 atm

14-18 Propylene Oxide S = 2
log H(mm Hg) = 8.244 - 1114/(t°C + 232)
Activity coefficient (35° to 100°C) = 15.3

15-1 Methyl Mercaptan S = 1
log H(mm Hg) = 7.032 - 1016/(t°C + 239) + 604/(t°C + 273)
log (Activitycoefficient)= 604/(t°C + 273)

15-2 Dodecyl Benzenesulfonic Acid S = 6
Ionizes in water, not strippable

15-3 Cyclohexane S = 1

log H(mm Hg) = 6.84 - 1202/(t°C + 223) + 1442/(t°C + 273)

log (Activitycoefficient)= 1442/(t°C + 273)

15-4 Isophorone S = 4

log H(mm Hg) = 7.99 - 1672/(t°C + 273)

log (Activity coefficient) = 836(t°C + 273)

15-5 Strychnine S = 6

No data

15-6 2,3,7,8-Tetrachlordibenzo-p-dioxin S = 3-P

No data

15-7 Zinc Phenol Sulfonate S = 6

Non-volatile, not strippable

APPENDIX F

SYMBOLS

A partial list of symbols, their common units, and an indication of where the symbol is first used or defined is given below. Symbols used once only, or in a limited context, are defined where used and not listed here. Cost symbols, defined in Section 4, are specifically not listed here. In general, any consistent system of units may be used. Some correlations are for specific units, however, and these are specified in context.

a_p	Superficial surface area, ft^2/ft^3	Table 3-4
A	Stripping medium rate, moles/hr	Eq. (3.1)
A,B,C	Antoine coefficients	Eq. (2.15)
B	Bottoms rate, moles/L	Eq. (3.1)
C_f	Packing factor, ft^{-1}	Table 3-4
d	Tower diameter	Eq. (3.29)
D	Diffusivity, ft^2/hr or m^2/s	
D	Overhead liquid rate, moles/hr	Fig. 3-2
E_A	Murphree efficiency corrected for entrainment	Table A-1

E_M	Murphree tray efficiency	Sec. 2.4.1
E_o	Overall tower efficiency	Sec. 2.4.1
E_p	Point efficiency	Sec. 2.4.1
f	Fractional removal efficiency	Eq. (3.3)
f	Fugacity, atm	Eq. (2.1)
F	Feed rate, moles/hr	Eq. (3.1)
g_c	Gravitational conversion constant, $4.18 \times 10^8\ lb_m.ft/(lb_f.hr^2)$	
G	Gas or vapor rate, lb/hr	Sec. 3.2.3
G'	Gas or vapor flux, $lb/(hr.ft^2)$	Sec. 3.2.3
h	Liquid enthalpy, cals/mole	Eq. (3.22)
H	Henry's Law constant, atm	Eq. (2.11)
H	Humidity, moles water/mole dry air	Eq. (3.25)
HTU	Height of a transfer unit, ft	Eq. (3.30)
i	Air enthalpy, cals/mole dry air	Eq. (3.23)
K	Vapor-liquid equilibrium constant	Eq. (2.11)
L	Total height of tower, ft	
L	Liquid rate, moles/hr and mass/hr	Fig. 3-2
n	Tray or stage number	

N	Total number of moles present	Eq. (2.2)
N	Total number of stages	
NTU	Number of transfer units	Eq. (3.30)
p	Partial pressure, atm	Eq. (2.3)
p^*	Vapor pressure of pure substance, atm	Eq. (2.8)
P	Total pressure, atm	Eq. (2.3)
R	Reflux ratio	Table 3-1
S	Stripping factor KV/L	Eq. (3.12)
t	Temperature, °F, °C	
T	Temperature, Kelvin	
U	Overall heat transfer coefficient	Table 3
V	Vapor rate, moles/hr	Eq. (3.1)
v	Vapor velocity in tower, ft/s	Eq. (3.26)
x	Mole fraction in liquid phase	Eq. (2.6)
y	Mole fraction in vapor phase	Eq. (2.2)
Y	Mole ratio in vapor	Eq. (3.19)
Z	Height of packing	Eq. (3.30)
ϵ	Packing voidage	Table 2-1

μ	Viscosity, $lb_m/(hr.ft)$	
ρ	Density, lb/ft^3	
γ	Activity coefficient	Eq. (2.7)
σ	Surface tension, dyne/cm	

Subscripts

B	Bottoms
F	Feed
G	Gas or vapor phase
i, 1, 2	Species
L	Liquid phase
m	Maximum
n, 1, 2	Stage number
t	Top of tower
V	Vapor phase